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Evaluation of the Limestone Dual Alkali Prototype System at Plant Scholz: Final Report

Thyssen-CEA Environmental Systems, Inc.
New York

Prepared for

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EVALUATION OF THE LIMESTONE DUAL ALKALI PROTOTYPE SYSTEM AT PLANT SCHOLZ: FINAL REPORT

by

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16. ABSTRACT

The report gives results of a 2-month test (February/March 1981) of the limestone dual alkali process at an existing 20 MW prototype facility at Gulf Power Company's Scholz Steam Plant. The project was intended to evaluate the technical feasibility of the process at a prototype scale and to develop sufficient technical information leading to the implementation of the process at a full commercial scale. Due to budgetary limitations, the test was reduced from 6 to 2 months. Although the technology appears to be technically feasible, further testing is necessary before a conclusive evaluation is made; some refinement of the process is still needed. Excellent SO_2 removal efficiencies (over 95%) were achieved; limestone utilizations were also high, over 97%. The solids properties ranged from excellent to poor. The generation of poor solids was the most significant process problem. The solids content in the filter cake ranged from 35 to 45% (below the anticipated 55%). The soda ash consumption of 0.29 moles of Na_2CO_3/mole of SO_2 removed far exceeded the design consumption of 0.04 moles. Much of this consumption was due to leaks and other liquor losses in the system.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Pollution Control	13B
Coal	Stationary Sources	21D
Combustion	Dual Alkali Process	21B
Desulfurization		07A,07D
Limestone		08G
Scrubbers		13I

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ABSTRACT

The limestone dual alkali process developed by Thyssen-CEA Environmental Systems, Inc. (until late 1980 under Combustion Equipment Associates, Inc. ownership) and Arthur D. Little, Inc. was tested at an existing 20 MW prototype facility at Gulf Power Company's Scholz Steam Plant. The intent of this project was to evaluate the technical feasibility of the process at a prototype scale and to develop sufficient technical information leading to the implementation of the process at a full, commercial scale.

Due to budgetary considerations, the testing period was reduced from six to two months. This report presents the results obtained during these two months of testing--February and March 1981. Excellent SO_2 removal efficiencies in excess of 95% were achieved; limestone utilizations were also high, over 97%. The solids properties ranged from excellent to poor. The generation of poor solids was the most signficant process problem. The solids content in the filter cake typically ranged between 35 and 45%, which was below the anticipated 55%. The soda ash consumption of 0.29 moles of Na_2CO_3/mole of SO_2 removed far exceeded the design consumption of 0.04 moles of Na_2CO_3/mole of SO_2. A substantial amount of this consumption was due to leaks and other liquor losses in the system. The mechanical performance of the equipment, recommissioned after three years of inactivity, was poor and contributed to the above problems.

Although the technology appears to be technically feasible, further testing is necessary before a conclusive evaluation is made, as some refinement of the process is still needed.

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APPLICABLE CONVERSION FACTORS

ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in	2.54 centimeters
1 in^2	6.452 $centimeters^2$
1 in^3	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal	3.7853 liters
1 Btu	252 calories

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I. SUMMARY

A. OBJECTIVES

The purpose of this test program was to evaluate the performance, at a prototype scale, of the limestone dual alkali process developed by Thyssen-CEA Environmental Systems, Inc. (TESI) (initially Combustion Equipment Associates, Inc. (CEA)) and Arthur D. Little, Inc. (ADL). The overall intent of this project, however, was not only to demonstrate the technical feasibility of the process at a prototype level but to supply technical and cost information related to the implementation of the process at a full, commercial scale.

The test program was conducted at an existing dual alkali facility at Gulf Power Company's (GPC) Scholz Steam Plant near Sneads, Florida. The 20 MW lime-based dual alkali system was modified and converted for operation with limestone. Following its conversion, the system underwent an evaluation of its performance with regard to:

- SO_2 removal capabilities;
- Raw materials and energy requirements;
- Quality of waste material generated; and
- Reliability and ease of operation of the system.

Originally, the formal testing period of the system was to last six months; this period was, however, reduced to two months due to budgetary considerations.

Parallel to this EPA program, the Electric Power Research Institute (EPRI) sponsored a program to study and evaluate the landfill disposal of the waste cake generated by the limestone dual alkali system.

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B. OPERATING HISTORY

The modification and conversion of the existing system to a limestone-based dual alkali system was completed in March 1980. Process operations were initiated in August 1980. In the interim, the system was operated mechanically in order to prevent mechanical deterioration of the equipment. The delay was caused by equipment delivery problems encountered in the installation of the EPRI waste disposal system. As the expenses associated with the mechanical operation continued to accumulate, it was decided to start up the system in August even though the EPRI system was not completely installed. Formal testing of the dual alkali system did not start until February 1981. During this period--August 20, 1980 to the end of January--the system was operated for about 57 days and was down for about 107 days. Of the downtime in this initial startup and break-in testing period, 43% was attributable to mechanical problems (e.g., leak in thickener tank), 20% to process problems (e.g., generation of solids with poor settling characteristics), 26% to the time needed to resolve CEA's bankruptcy-related issues[1], 7% to lack of raw materials (due to reluctant suppliers in light of the bankruptcy declaration), and 4% to miscellaneous causes (e.g., boiler outage).

Testing of the system began on Feburary 2, 1981 and extended until March 28, 1981. During this period, the system was operated for 925 hours (38.5 days) or 71.4% of the time, and it recorded one completely uninterrupted period of operation of 431 hours (18 days).

The outages in the testing period were due primarily to mechanical problems which accounted for 63% of the downtime, and to process problems which accounted for 30% of the downtime.

C. SYSTEM PERFORMANCE

Given the limited testing performed, due to the reduction in the testing period from six to two months, a final and conclusive evaluation of the technology, leading to its full-scale commercialization cannot be made at this time. Nevertheless, a number of very encouraging qualities of this technology have been clearly identified. Problem areas that will require further refinement have been identified as well.

1. General System Description

The limestone dual alkali system involved the absorption of SO_2 using an aqueous solution of alkaline sodium salts. Regeneration of the spent absorbent solution was accomplished using limestone which produced a solid waste of calcium-sulfur salts. The system was intended to operate in a closed loop, the only waste material being a washed filter cake. The system consisted of basically four process subsections: flue gas contacting (SO_2 absorption); absorbent solution regeneration and formation of waste solids; waste solids dewatering (thickening and filtration); and raw materials storage and feed preparation (for soda ash and limestone).

2. SO_2 Removal

The SO_2 removal capability was excellent. During the testing period, the system averaged an SO_2 removal efficiency of 95.8%. SO_2 removal efficiencies were easily controlled by simply adjusting the pH of the scrubber bleed liquor. Inlet SO_2 concentrations during this period ranged from 1460 to 3240 ppm and averaged at slightly over 2000 ppm.

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[1] On October 20, 1980 Combustion Equipment Associates, Inc. filed a petition for financial reorganization pursuant to Chapter 11 of the U.S. Bankruptcy Code. Thyssen-CEA Environmental Systems, Inc. (TESI) bought the CEA Air Pollution Division and administered this EPA contract starting in early December 1980, assuming financial responsibilities retroactive to October 21, 1980. The EPA contract, subject of this project, was formally assigned to TESI in April 1981. CEA is referenced in this report to maintain the historic integrity.

3. Limestone Utilization

The limestone utilization by the dual alkali system was very good. During the test period, utilizations in the reactor train effluent ranged from 85 to 95% of the available $CaCO_3$ in the raw limestone. As the reaction with limestone continued in the dewatering system, the final system utilizations, determined from chemical analyses of the filter cake, ranged from 93 to 100%, averaging at 97.5%.

4. Oxidation and Sulfate Precipitation

Approximately 18% of the SO_2 removed by the scrubbing solution was oxidized to sulfate as it passed through the system. As anticipated, most of this oxidation, as much as 90%, took place in the absorber/scrubber section. The capabilities of the system to co-precipitate calcium sulfate along with calcium sulfite were adequate to remove the sulfate formed. The soluble sulfate concentration in the system achieved a steady level around 1 M. Although liquor losses (due to leaks, spills, and purges to maintain the liquor volume balance in the system) contributed to the removal of sulfate from the system, the sulfate concentration in a tight, closed loop operation is not expected to rise above 1.2 M, which appears to be well within the sulfate co-precipitating capabilities of the system.

5. Waste Solids

The generation of solids with good settling characteristics was the most significant process limitation encountered at Scholz. It essentially accounted for all of the process-related outages. Throughout the months of December and February, solids with excellent settling characteristics were generated. In contrast, poor solids were generated in January. The good settling solids were agglomerates, roughly spherical in shape, whereas the poor solids were fine, needle shaped solids. These fine solids, whenever carried over in the thickener overflow in noticeable amounts (>1000 ppm), promoted the formation in the reactors of more fine and difficult to settle solids. While on some occasions, the formation of these fine solids was stopped by process changes; in others it simply continued to deteriorate forcing interruptions in the operation of the system. Although some mechanical problems intensified the difficulties in the settling of solids, the fundamental physical and chemical reasons for the formation of poor settling solids is not clearly understood; and therefore requires further investigation.

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The solids appeared, in general, to have good dewatering characteristics; nevertheless, the typical solids content in the cake ranged from 35 to 45 wt.%, far below the anticipated 55 wt.%. Severe mechanical problems with the filter and associated pumps and piping were the major factors in this shortcoming. Closely associated with this problem was the high sodium losses in the waste filter cake.

It appears that both of these problems--low solids content of the cake and high sodium losses--are not problems inherent to the limestone dual alkali technology, but were rather caused by the mechanical condition of the equipment used and by operational problems specific to this test program.

6. Soda Ash Consumption

The soda ash consumption was excessive, amounting to about 0.29 mole of Na_2CO_3/mole of SO_2 removed. Soda ash fed to the system is meant to make up for normal sodium losses in the liquor entrained in the filter cake. At Scholz, however, not only were the sodium losses high in the filter cake; but a substantial amount of sodium was lost through leaks, spills, and liquor purges needed to maintain the liquor volume balance in the system. (Large amounts of seal water were needed to keep worn out pumps in operation which, coupled with heavy rains, were very taxing on the limited surge capacity available.)

Again, this excessive soda ash consumption problem does not appear to be inherent to the limestone dual alkali technology.

7. Power Consumption

The power consumption ranged from 2.5% (0.53 MW) at flue gas rates equivalent to a boiler load of 21 MW, to 5.3% (0.42 MW) at an equivalent load of 8 MW. In a typical limestone dual alkali application, where flue gas is taken from an electrostatic precipitator and the use of a venturi is not required, one would expect the power consumption to be much lower--more in the range of 1 to 1.5% of the power generated at full boiler load.

8. Process Operability

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Process operability, unlike equipment or mechanical operability, refers to the ease with which the system can be operated and controlled, to the ability of the system to adequately respond to varying conditions, and to the ability of the system to tolerate upsets in process chemistry due to mechanical problems or operator oversight.

In general, the process operability of the system was good after the first two or three days of stable operation. It was during these initial days, following any restart of the system, that problems with process operability were encountered. All of these related to the generation of solids with poor settling characteristics.

Once the system reached a stable operation, such as in the month of December or February, the process operability was rather good. The system was able to easily accommodate variations in inlet SO_2 concentrations of as much as 500 ppm, by simply adjusting the feed forward rate of regenerated solution to the absorber/scrubber in order to maintain a constant scrubber bleed pH. Variations in boiler load and thus in the amount of gas processed were also accommodated in the same fashion.

Upsets to the process chemistry were also handled well by the system. These upsets included the carryover of fly ash in the flue gas due to a malfunction in the precipitators, the gross overfeeding of limestone due to operator's oversight, and the occasional limestone and soda ash outages lasting from 1 to 5 hours.

9. Mechanical Performance

The mechanical performance of the equipment and its associated instrumentation represented a major source of problems at Scholz. This is evidenced by the fact that 43% of the system downtime during the startup and break-in period and 59% of the downtime during the testing period were due to mechanical problems. Not only was the availability of the process affected by these mechanical problems, but the operating conditions that could be achieved by the system were affected as well. An example of these limitations was the inability of the thickener underflow pumps to handle slurry with more than 15 wt.% solids, which required the dilution of the 20 to 25% thickener underflow slurry. Although some of the problems were caused by the limited capacity of an existing piece of equipment, most of the mechanical problems were actually due to equipment failures. The age of the equipment and its condition, even after recommissioning, were, undoubtedly, primary causes of these failures.

D. CONCLUSIONS

The performance of the system as summarized above leads to the following conclusions:

- The limestone dual alkali process appears to be technically feasible; and

- Further testing is required to reinforce the above conclusion and to develop sufficient process information needed for full-scale commercialization purposes.

It is also recommended that laboratory or small pilot plant tests, to better understand the generation of solids with good settling characteristics, be performed prior to any further prototype scale testing.

II. INTRODUCTION

A. BACKGROUND

Thyssen-CEA Environmental Systems, Inc. (initially Combustion Equipment Associates, Inc. (CEA)) and Arthur D. Little, Inc. (ADL) developed, over the past few years, a dual alkali process for removing SO_2 from flue gas generated in coal-fired utility boilers. This process is based on the absorption of SO_2 in an alkaline sodium solution, followed by regeneration of the absorbing solution by reaction with a second alkali, calcium. These reactions generate insoluble calcium-sulfur salts which are discharged from the system as a moist cake.

The dual alkali process presents two significant advantages over conventional direct lime or limestone scrubbing technology. First, it uses a clear liquor, rather than a slurry, for scrubbing the flue gas. Second, the regeneration of the scrubbing solution and precipitation of waste solids takes place outside the absorber. Thus, the potential for scaling and plugging in the absorber is minimized; the formation of solids with good dewatering properties is more easily controlled; and high SO_2 removal efficiencies (>90%) are easily achieved with alkaline sodium scrubbing solutions by simply adjusting the scrubber operating pH.

The dual alkali process has been tested extensively at laboratory, pilot plant, and prototype levels using lime (calcium hydroxide) as the source of calcium for the regeneration reactions[1]. The performance of the process in these test programs prompted the federal Environmental Protection Agency (EPA) to select it for a demonstration plant. This lime-based demonstration plant has been installed on a 300 MW boiler at Louisville Gas and Electric's Cane Run Station. The system is currently undergoing a one-year test program.

Although the lime-based system is technically and economically viable, a source of calcium cheaper than lime would increase the economic attractiveness of the process. Limestone (calcium carbonate) was recognized early on as a potential source of cheaper calcium for the dual alkali process. Extensive testing of the use of limestone was undertaken by ADL at laboratory and pilot plant levels under funding both by EPA[1,2] and CEA. Successful pilot operations using limestone were achieved in 1977. In an extended run using Fredonia limestone, SO_2 removal efficiencies greater than 90% were attained with a limestone utilization of 99+%; the waste filter cake contained 55 to 60% insoluble solids. In other testing performed with different limestones, the reactivity and efficiency of utilization of limestone was found to be strongly dependent on its particle size distribution--the smaller the particles, the better the performance of the limestone.

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[1] LaMantia, C.R., et al., "Final Report: Dual Alkali Test and Evaluation Program," Volume I-III, EPA-600/7-77-050a,b,c, May 1977.

[2] Oberholtzer, J.E., et al., "Laboratory Study of Limestone Regeneration in Dual Alkali Systems," EPA-600/7-77-064, July 1977.

The successful performance of the system during the laboratory and pilot plant test programs revealed the potential of the limestone-based dual alkali process for full-scale applications. This project was aimed at confirming this potential as well as providing needed information for its full-scale application.

B. PROJECT OBJECTIVES

The objectives of this program were to demonstrate the technical and economic feasibility of the limestone dual alkali technology for application in flue gas desulfurization (FGD) of high sulfur, coal-fired boilers.

Under this program, the existing lime dual alkali system at Gulf Power Company's (GPC) Scholz Steam Plant was modified and converted for the exclusive use of limestone as the regenerating material.

After modification and conversion to a limestone-based process, the system was to undergo a test program to evaluate its performance with respect to:

- Capabilities for removal of SO_2.
- Raw materials and energy requirements.
- Quality of the waste material generated.
- Reliability and ease of operation of the system.

Furthermore, capital and operating costs were to be estimated for full-scale utility applications of the limestone dual alkali FGD technology.

Parallel to this EPA program, the Electric Power Research Institute (EPRI) sponsored a program to study and evaluate the landfill disposal of the waste cake generated by the limestone dual alkali system. Under the EPRI program the FGD waste cake was to be treated with fly ash and lime prior to its disposal. The program itself was to consist of: monitoring the waste disposal system to characterize its operation; analyzing the physical and chemical properties of the treated waste during and after disposal operations; and establishing the potential for revegetation of the waste disposal area.

Originally, the EPA test program of the limestone-based dual alkali system[3] was scheduled to last six months exclusive of startup procedures and it was to be conducted in two periods. During the first period, approximately two months, a baseline for the operation of the system was to be established. The second period, approximately four months was to consist of variational testing of the system. However, because of budgetary constraints, due to actual costs exceeding those budgeted, delays due to the need to coordinate with the EPRI program, and additional delays caused by CEA's declaration of bankruptcy, the testing period was reduced from six to two months--February and March 1981.

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[3] Valencia, J.A., et al., "Evaluation of the Limestone Dual Alkali System at the Scholz Plant - System Design and Program Plan," Report to EPA, Contract 68-02-3128, March 1981.

III. THE SCHOLZ PROTOTYPE SYSTEM

A. LIMESTONE DUAL ALKALI TECHNOLOGY

The limestone dual alkali technology consists of four distinct operations: SO_2 absorption; absorbent regeneration; waste solids dewatering; and raw materials storage and feed preparation.

Absorption of SO_2 is accomplished by contacting flue gas with a sodium sulfite/bisulfite solution. The sulfite reacts with the SO_2 producing additional bisulfite according to the following overall reaction:

$$SO_3^{=} + SO_2 + H_2O \rightarrow 2\ HSO_3^{-}$$

During absorption, and to a lesser extent throughout the remainder of the process, some sulfite is oxidized to sulfate:

$$SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=}$$

The level of oxidation experienced during absorption is generally a function of the scrubber configuration, oxygen content of the flue gas, and the scrubber operating temperature. At excess oxygen concentrations normally encountered in utility power plant operations, the level of oxidation is expected to be on the order of 5 to 10% of the sulfur dioxide removed (for medium and high sulfur coal applications).

The spent scrubbing solution is reacted with limestone to regenerate the absorbent. The reaction precipitates mixed calcium sulfite and sulfate solids, resulting in a slurry containing up to 5 wt.% insoluble solids. The regeneration process involves basically the following overall reactions:

$$2\ HSO_3^{-} + CaCO_3 \rightarrow SO_3^{=} + CaSO_3 \cdot 1/2\ H_2O\downarrow + 1/2\ H_2O + CO_2\uparrow$$

$$2\ HSO_3^{-} + Na_2SO_4 + CaCO_3 \rightarrow 2\ SO_3 + CaSO_4 \cdot 1/2\ H_2O\downarrow + 1/2\ H_2O + CO_2\uparrow$$

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Following regeneration, the insoluble calcium-sulfur salts are separated from the regenerated liquor. After settling in a thickener, the insoluble salts are vacuum filtered prior to being discharged as a moist cake. The clarified liquor from the thickener is returned to the scrubber, closing the process liquor loop.

The filter cake is washed to recover most of the soluble sodium salts in the mother liquor. A small portion of the sodium salts, however, remains occluded in the cake and is lost. Thus, some sodium carbonate must be added to the system to make up for these sodium losses. Typically, the Na_2CO_3 makeup should amount to less than 5 mole % of the SO_2 removed.

The amount of limestone added to the system for absorbent regeneration is reduced slightly due to the soda ash makeup. Under normal conditions, the limestone feed stoichiometry will be slightly less than one mole of $CaCO_3$ per mole of SO_2 removed.

B. DESCRIPTION OF THE SCHOLZ PLANT

The Scholz Steam Plant, located in Sneads, Florida, is the property of Gulf Power Company[1]. The facilities in this plant, as they relate to the present project, are best described in three sections: the boiler, the limestone dual alkali system, and the waste disposal facilities.

1. Boiler

The dual alkali system took flue gas from Unit No. 1, a 40 MW nominal capacity (47 MW peak capacity) Babcock and Wilcox pulverized-coal-fired power boiler. The flue gas from the boiler passed through a high efficiency, sectionalized electrostatic precipitator designed to remove up to 99.5% of the particulate matter. Part of the gas discharged from the precipitator was directed to the limestone dual alkali system, which was designed to handle a flue gas flow rate equivalent to a 20 MW boiler load. The remaining gas was sent directly to the stack.

2. Limestone Dual Alkali System

The limestone dual alkali system at Scholz was a modification of a lime dual alkali system previously installed and operated during 1975 and 1976 for a test program sponsored by The Southern Company and jointly funded with EPA. The modified dual alkali system consisted of four sections as discussed previously: absorption; regeneration; waste solids dewatering; and raw materials storage and feed preparation. A process flow diagram for the limestone dual alkali system at Scholz is given in Figure III-1.

a. Absorption

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The absorption system, shown in Figure III-2, consisted of a venturi scrubber, followed by an absorption tower. A sulfur dioxide injection system was provided to increase, when necessary, the SO_2 concentration in the flue gas entering the dual alkali system by 200 ppm or more. The SO_2 was injected upstream of the booster fan, before the gas entered the venturi scrubber.

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[1] Gulf Power Company is part of The Southern Company, an electrical utility holding company operating in the Southeast. It includes Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, and Southern Company Services, Inc.

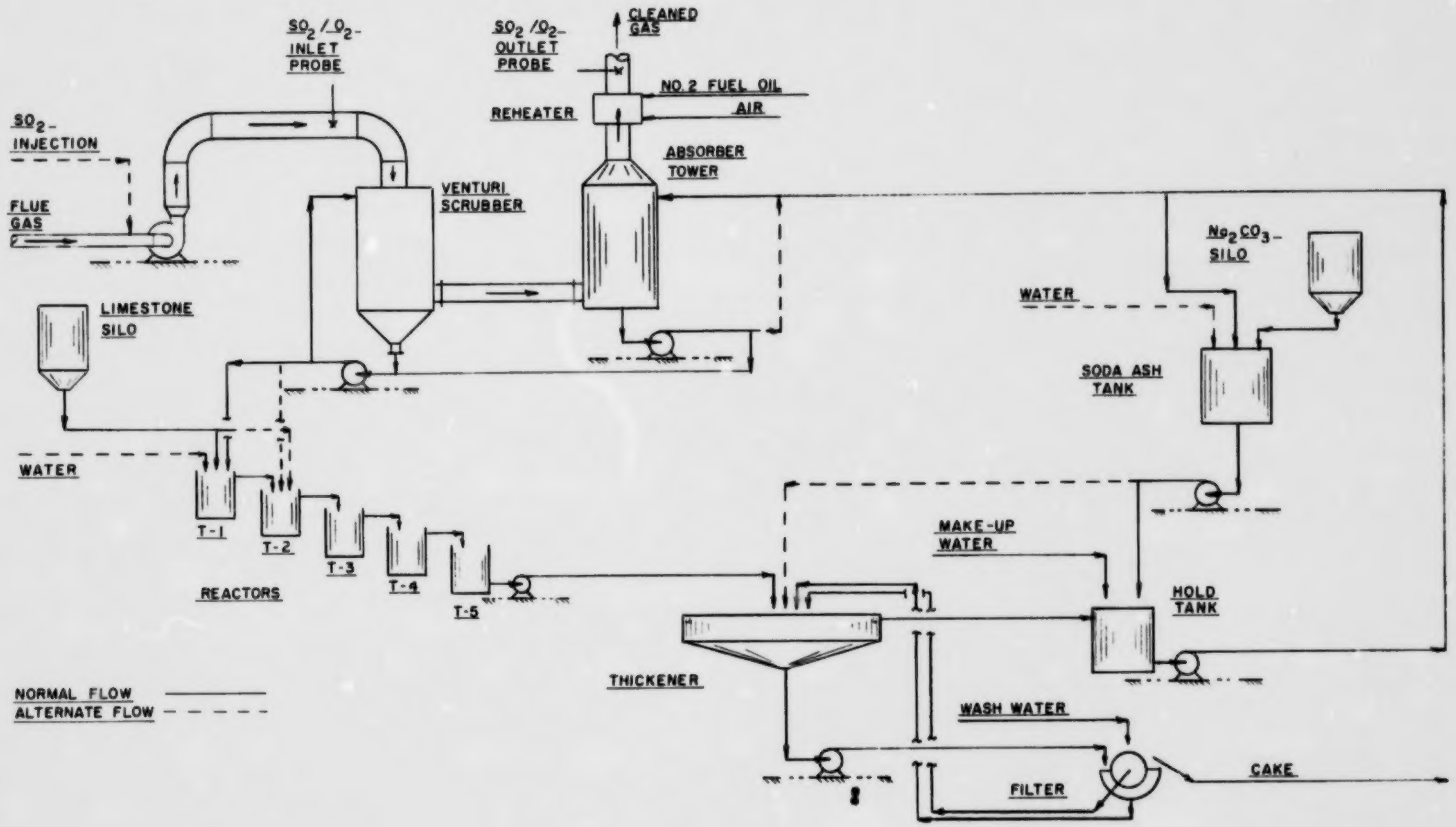
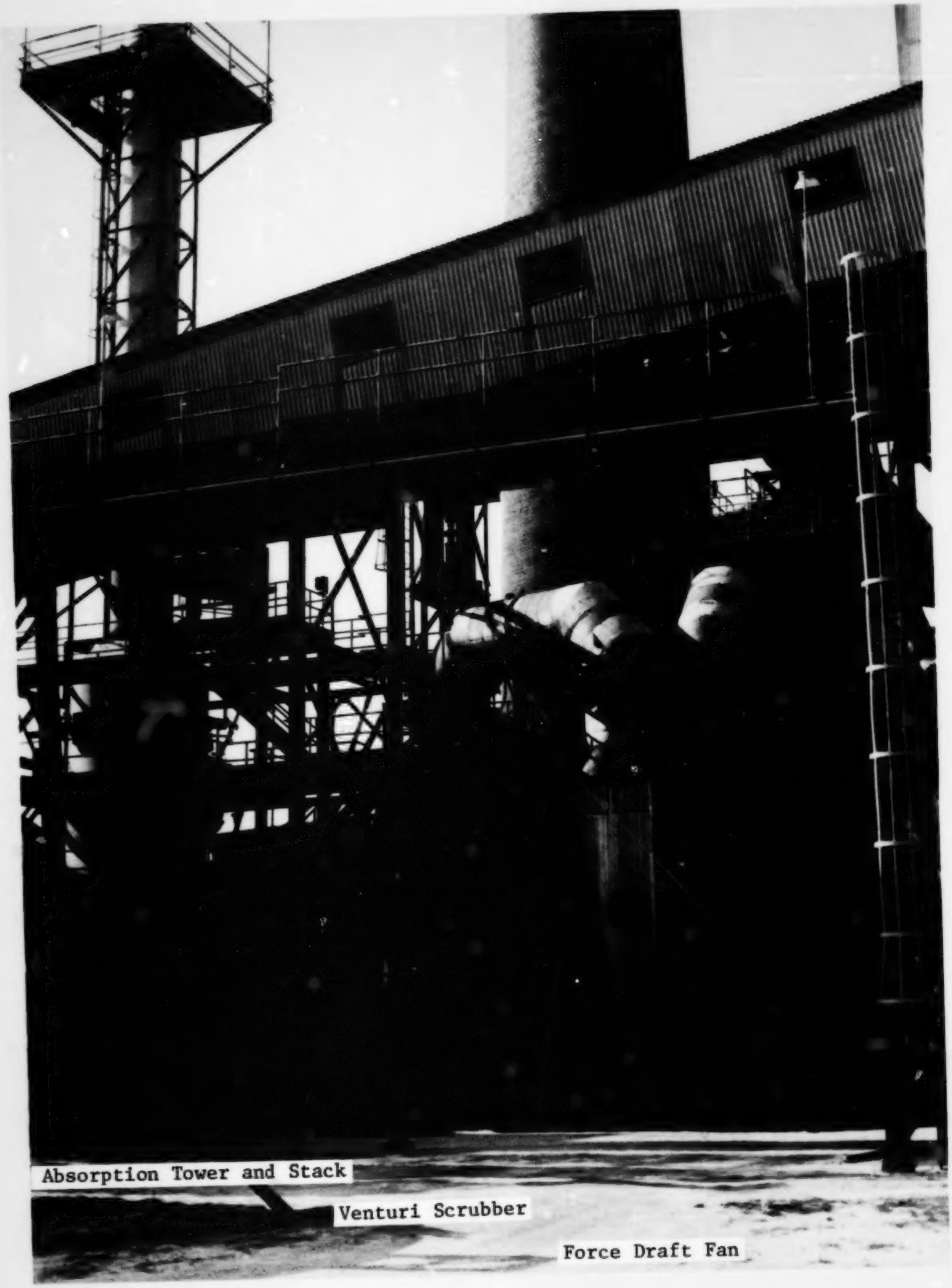

SO_2/O_2-
INLET
PROBE
SO_2/O_2-
OUTLET
PROBE
CLEANED
GAS
NO. 2 FUEL OIL
AIR
REHEATER
ABSORBER
TOWER
SO_2-
INJECTION
FLUE
GAS
VENTURI
SCRUBBER
LIMESTONE
SILO
Na_2CO_3-
SILO
WATER
SODA ASH
TANK
WATER
T-1
T-2
T-3
T-4
T-5
REACTORS
MAKE-UP
WATER
HOLD
TANK
THICKENER
NORMAL FLOW
ALTERNATE FLOW
WASH WATER
CAKE
FILTER

Figure III-1: Limestone Dual Alkali System at Scholz - Process Flow Diagram



Absorption Tower and Stack
Venturi Scrubber
Force Draft Fan

Figure III-2: SO_2 Absorption System

The variable throat, plumb-bob type venturi scrubber was designed for particulate removal and/or SO_2 absorption. Typically, a dual alkali system, taking flue gas from a particulate removal device (e.g., electrostatic precipitator) would not require a venturi scrubber. Such being the case at Scholz, the system in fact did not require the venturi scrubber. Rather than remove it, the venturi scrubber was used primarily for quenching and saturating the flue gas; these operations, which also contribute to the SO_2 removal, would normally be performed in the absorption tower.

A booster fan directed the SO_2-rich flue gas exiting the electrostatic precipitator into the scrubber, where it flowed downward over the wetted approach section and into the high velocity venturi throat. Recycled scrubbing liquor, at an L/G of 15 to 20 gallons/Macf, also entered the top of the scrubber through tangential pipe inlets and through a number of vertical bull nozzles equally spaced around the center of the venturi. After passing through the throat, the saturated gas and scrubbing liquor flowed downward through the internal downcomer. The gas made a 180° upward turn and passed to the absorption tower. The liquor was collected at the bottom of the scrubber in the internal recycle tank.

After being quenched and saturated with water, the partially desulfurized gas entered the bottom of the absorption tower. This tower was designed to operate either as a spray tower (with one to three sets of sprays) or as a tray tower (with one to four trays). For operation in the limestone dual alkali mode, the tower was operated with two trays. The bottom tray was equipped with a spray underneath to wet the bottom side of the tray. Gas passed upward through the trays and then through a de-entrainment separator. The de-entrainment separator was operated dry, without mist eliminator wash water.

Scrubbed flue gas leaving the tower was finally reheated by the injection of the hot gas from an oil-fired reheater before being discharged through the stack on top of the absorber.

10 Lines to End of Typing Area

The scrubbing solution, which was fed to the top tray of the absorber at an L/G of 2 to 3 gallons/Macf, flowed countercurrent to the gas through the tray system. This counterflow permitted high SO_2 transfer from gas to liquid phase. The liquor was collected at the bottom of the absorber in the internal recycle tank, which served as the supply for the liquor sprays beneath the bottom tray. Provisions were also made to recycle liquor from the internal tank to mix with fresh feed to maintain liquid loading on the trays at low boiler loads. As the demand for recycle liquor was small compared to the pump capacity, most of the recycle liquor was directed back to the internal recycle tank via a long bypass downcomer.

A bleed from the absorber liquor recycle line was sent forward on recycle tank level control to the scrubber recirculation loop where it was used to quench the gas and provide additional SO_2 removal. A bleed stream of spent liquor was drawn from the scrubber recycle line and was fed to the absorbent regeneration system at a rate controlled by the liquid level in the scrubber internal tank.

b. Absorbent Regeneration

Spent solution from the scrubber recirculation line was bled (on level control) to a multi-stage reactor system which was designed to have the capability of being operated in a number of configurations. The reactor system consisted of five separate tanks: the existing primary reactor (750 gallons) followed by four new equal-sized tanks (3400 gallons each) in series with a total design holdup time of approximately 100 minutes. The reactor system is shown in Figure III-3. The new tanks were identical in design, each being a cyclindrical, baffled vessel with a center-mounted agitator and with two overflow parts providing a capacity of either 2200 or 3400 gallons (except for the last tank that was equipped with a pump and operated on level control).

The new tanks were located adjacent to the existing reactor system for ease in interfacing with the limestone feed system and the existing primary reactor tank, and to allow for greater flexibility in the reactor system operation. The bleed from the scrubber system was piped such that it could be fed either to the existing first stage reactor (T-1) or to the first of the four new reactor tanks (T-2). Similarly, the modified limestone silo feed chute had the flexibility of feeding dry limestone to either of these two tanks. Thus, the modified reactor system was capable of being operated as either four or five reactors in series with dry limestone feed, or as four reactors in series with slurried limestone feed prepared in the existing first stage reactor. The capability also existed for varying the holdup time by using the low or high overflow parts. Due to the shortened testing period, however, it was operated in only one mode: five reactors in series with dry limestone feed to the first reactor. The first four reactors operated on overflow (high overflow for the three intermediate reactors), while the last reactor was on level control so that the effluent could be pumped to the thickener.

10 Lines
to End of
Typing Area

c. Waste Solids Dewatering

The dewatering system is shown in Figure III-4. Slurry from the reactor system was pumped to the centerwell of the existing 93,000 gallon slurry thickener. Clear liquor overflow from the thickener was collected in the existing thickener hold tank which acted as surge capacity for the absorbent liquor feed to the scrubber system. Process makeup water was added to this tank to make up for the total water lose due to evaporation in the system and moisture in the cake.

The underflow slurry was pumped through a recirculation loop around the thickener to prevent settling of the solids and plugging of these lines. The feed to the existing rotary drum vacuum filter was taken as a bleed from this recirculation line.

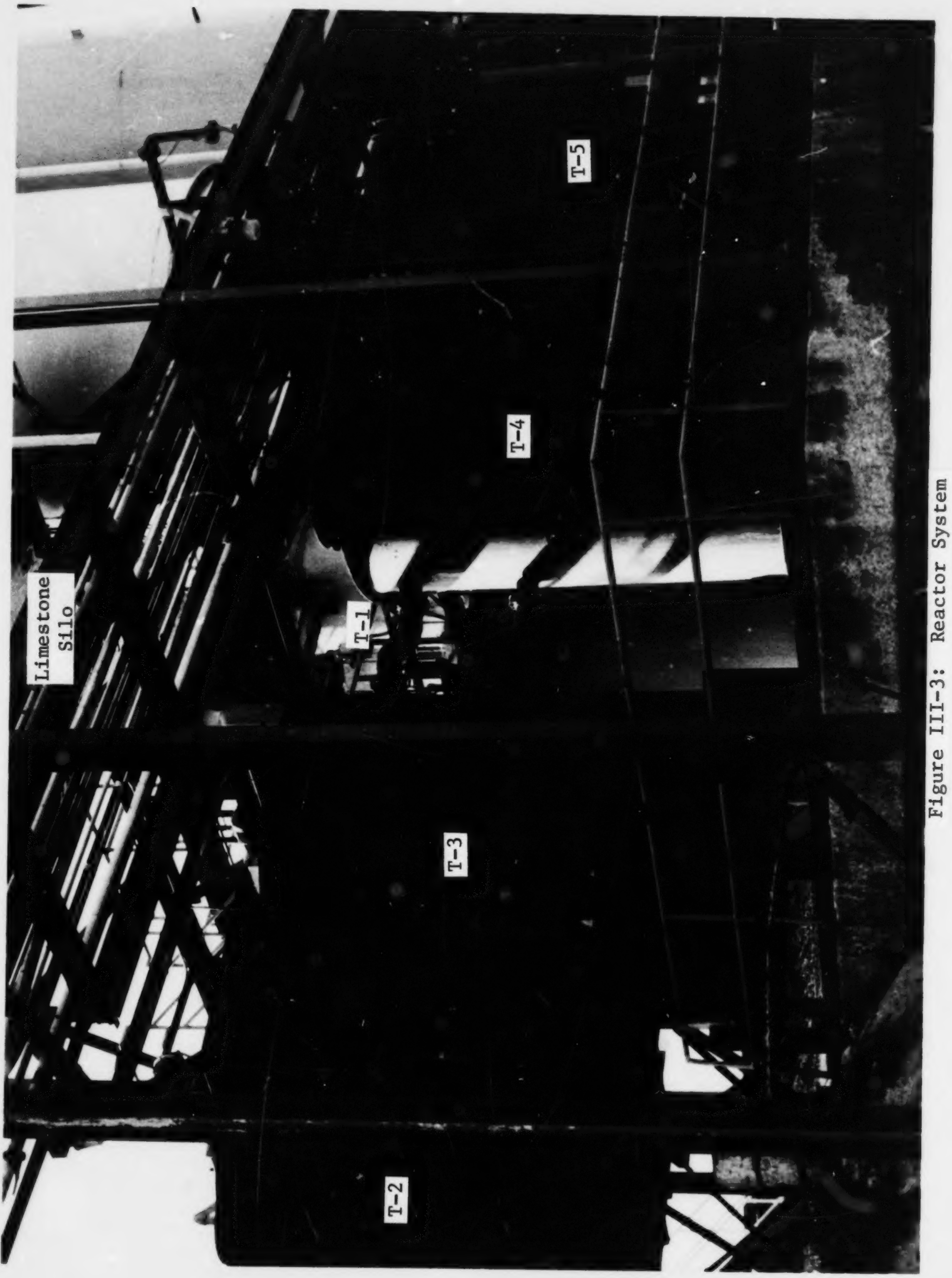

Limestone Silo
T-1
T-2
T-3
T-4
T-5

Figure III-3: Reactor System

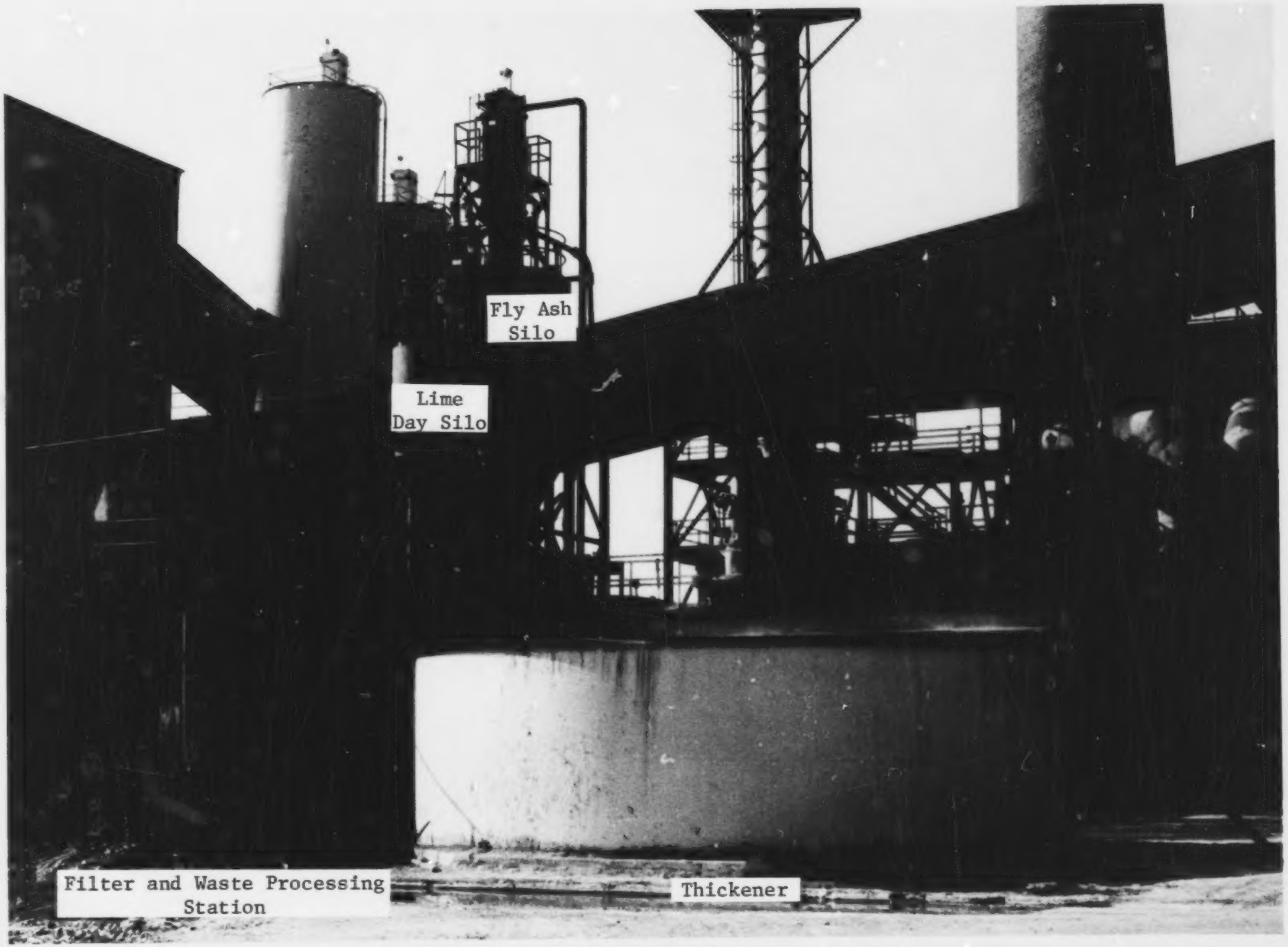

Fly Ash
Silo
Lime
Day Silo
Filter and Waste Processing
Station
Thickener

Figure III-4: Solids Dewatering System

The filter operated on overflow using an adjustable weir to set the proper slurry level in the filter tub. The overflow slurry was returned to the thickener. The solids cake was washed on the filter drum using two wash water spray banks. Solids from the filter cake were discharged to a weigh belt conveyor for transport to the waste processing system. The mixed filtrate and wash liquor from the filter were returned to the thickener.

d. Raw Materials Preparation

The raw materials preparation system is shown in Figure III-5. Two raw materials were required for operation of the system--limestone for absorbent regeneration and soda ash to make up for losses of sodium salts in the waste filter cake. (Lime was used exclusively for the EPRI waste disposal program.)

Finely ground limestone (97 to 99.9% through 325 mesh) was received, stored, and fed to the system in a dry form. As indicated above, the flexibility existed for feeding it as a slurry but that option was not exercised.

Dense soda ash was used to make up for sodium salt losses. The flexibility also existed to feed soda ash solution to the thickener, thickener hold tank, absorber recycle, or venturi scrubber recycle. It was normally fed to the thickener hold tank.

The soda ash solution was typically prepared using clarified liquor from the thickener hold tank. River water could also be used for preparing this solution but only when this additional water input did not upset the system volume balance. The makeup soda ash solution was fed back to the thickener hold tank to be mixed with the regenerated scrubbing solution and pumped forward to the absorber.

3. Waste Processing System

The waste processing system was installed and operated as part of a separate test program funded by EPRI. It is briefly discussed in this report to indicate its relation to the overall plant operation.

The moist waste cake discharged from the vacuum filter was mixed in a pug mill with fly ash and lime. Fly ash was added at approximately one part of ash per part of dry solids in the cake. Lime was added at a rate of 3 to 5 wt.% of the dry solids contained in the mix of cake and fly ash.

The mix discharged by the pug mill was loaded onto a truck and transported to a landfill test area. Mixed material not meeting specifications or untreated filter cake--when the waste processing system was unavailable or inoperative--were disposed of in a separate "off-spec" disposal area.

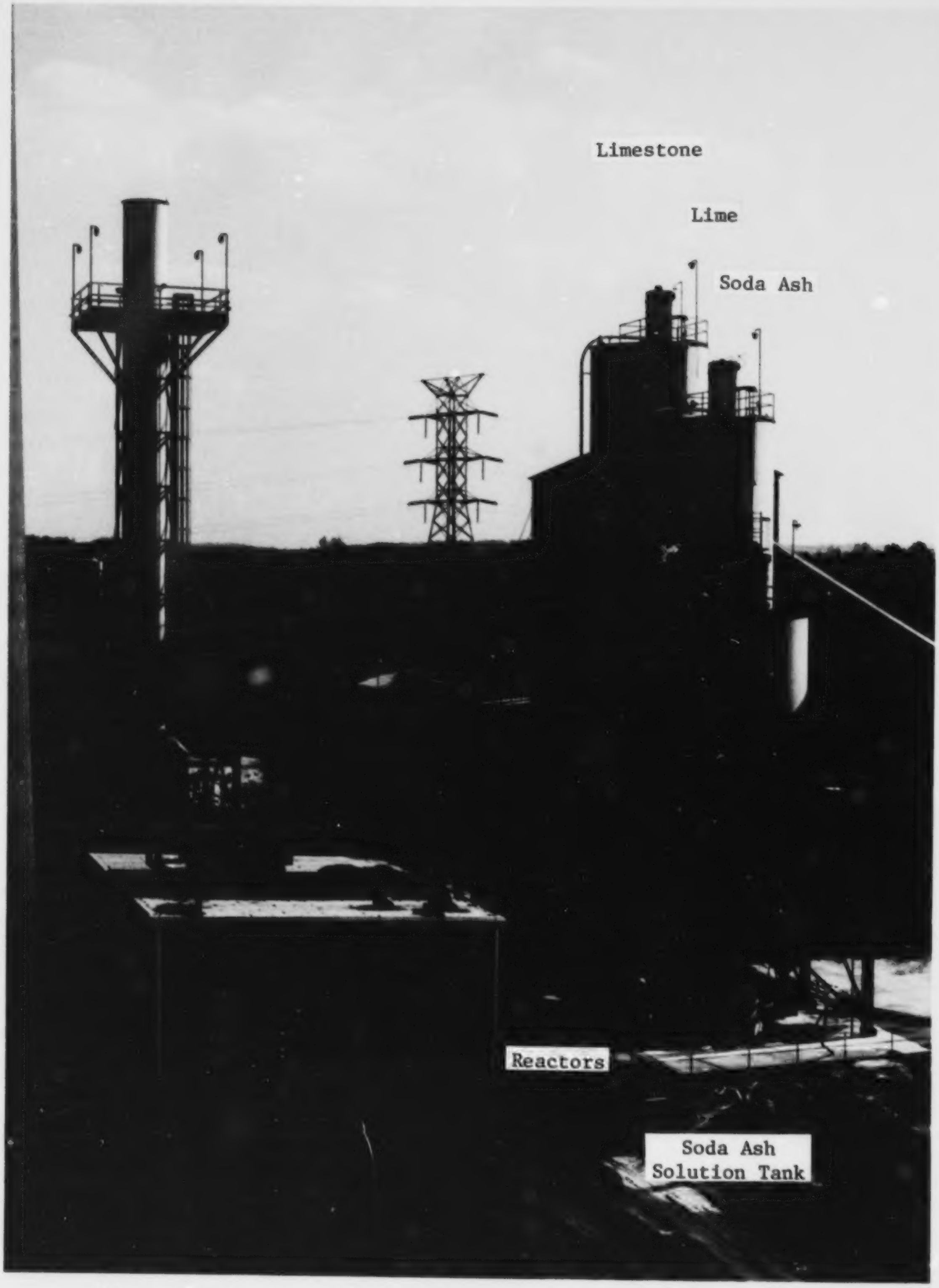

Limestone
Lime
Soda Ash
Reactors
Soda Ash Solution Tank

Figure III-5: Raw Materials Storage and Preparation System

C. SYSTEM DESIGN BASIS

The design of the limestone dual alkali system at Scholz assumed the firing of the boiler with coal containing 2.9 to 3.4 wt.% sulfur. The corresponding SO_2 levels in the boiler flue gas were estimated to be on the order of 1900 to 2300 ppm. As it was desirable to test the high end of this range and even higher inlet SO_2 concentrations, an SO_2 injection system was provided. Additional SO_2 could then be injected into the flue gas system before it entered the scrubber. The limestone dual alkali system was, therefore, designed to handle a normal operation as well as a maximum load operation with regard to SO_2 loading. The normal operation was based on a gas flow of 40,000 dscfm and an inlet SO_2 concentration of 2200 ppm (dry). The maximum operation was based on a gas flow of 45,000 dscfm and an inlet SO_2 concentration of 2650 ppm (dry). Additional details on the design bases are given in Table III-1 for the normal operation and in Table III-2 for the maximum operation.

10 Lines to End of Typing Area

TABLE III-1

DESIGN BASIS - NORMAL OPERATION

Inlet Gas:	
Flow Rate	40,000 scfm (dry)
SO_2	14.9 lb/min. (2,200 ppm dry)
Cl^-	0.12 lb/min.
O_2	5 vol.%
Moisture	6.0 wt.%
Removal Efficiency:	
SO_2	95%
Cl^-	100%
Absorber/Scrubber Feed:	
Absorber Tray Feed	2.4 gallons/1000 acf
Na^+ associated with	
$SO_3^=$	1.02 M
HSO_3^-	0.68 M
Scrubber Recycle	16 gallons/1000 acf
Oxidation and Evaporation:	
Oxidation	7 mole % of ΔSO_2
Evaporation in Absorber	22 gpm
Soda Ash and Limestone Feed:	
Soda Ash Feed Rate	0.07 mole Na^+/mole ΔSO_2
Limestone Purity	95 wt.%
Limestone Feed Rate	0.986 mole available $CaCO_3$/mole ΔSO_2
Waste Solids:	
Wash Ratio	4 displacement washes
Insoluble Solids	55 wt.%

10 Lines to End of Typing Area

TABLE III-2

DESIGN BASIS - MAXIMUM LOAD OPERATION

Inlet Gas:	
Flow Rate	45,000 scfm (dry)
SO_2	20.2 lbs/min. (2,650 ppm dry)
Cl^-	0.12 lbs/min.
O_2	5 vol.%
Moisture	6.0 wt.%
Removal Efficiency:	
SO_2	95%
Cl^-	100%
Absorber/Scrubber Feed:	
Absorber Tray Feed	2.6 gallons/1000 acf
Na^+ associated with	
$SO_3^=$	1.02 M
HSO_3^-	0.68 M
Scrubber Recycle	15 gallons/1000 acf
Oxidation and Evaporation:	
Oxidation	10 mole % of ΔSO_2
Evaporation in Absorber	24.7 gpm
Soda Ash and Limestone Feed:	
Soda Ash Feed Rate	0.081 moles Na^+/mole ΔSO_2
Limestone Purity	90%
Limestone Feed Rate	1.010 moles available $CaCO_3$/mole ΔSO_2
Waste Solids:	
Wash Ratio	4 displacement washes
Insoluble solids	55 wt.%

10 Lines to End of Typing Area

IV. OPERATING HISTORY

The tasks of designing the system and planning the test program began in September 1978 immediately after EPA awarded the contract for this project. The recommission and conversion of the existing system to a limestone dual alkali system began in March 1979 and was completed by March 1980. Process startup and debugging operations took place between August 1980 and January 1981. Testing of the system was performed in the months of February and March 1981. The significant events during the operating periods are discussed below; operating charts for the dual alkali system and the boiler, which provide details on their operation and outage history, are given in Appendix A.

A. STARTUP

Mechanical operation of the dual alkali system started on April 28, 1980. This operation was primarily intended to prevent the mechanical deterioration of the equipment while awaiting the delivery and installation of the EPRI waste disposal system. As the expenses associated with this mechanical operation continued to accumulate, EPA decided, near the end of July 1980, to proceed with the process operations even though the waste disposal system was not completely installed. CEA's estimates at the time indicated that the remaining balance in the budget would permit only 4.7 months of operation; 2 months for startup and 2.7 months for testing.

The startup and break-in testing phase commenced on August 13, 1980 and ended on January 31, 1981. Flue gas was first admitted to the system on August 20, 1980. Between this date and end of this initial phase, the system processed flue gas for 1,380 hours or 35% of the time. Of the downtime in this initial phase, 43% is accounted for by mechanical problems (e.g., leak in the thickener tank), 20% by process problems (e.g., generation of solids with poor settling behavior), 26% by CEA's bankruptcy-related interruptions, 7% by lack of raw materials, and 4% by miscellaneous causes (e.g., boiler outage).

Round-the-clock operations began on August 13, 1980. After final checkout procedures were completed, flue gas was admitted to the system on August 20, 1980. Filling of the system was accomplished by feeding fresh soda ash solution to the absorber; after removing SO_2, this solution had a sulfite/bisulfite concentration near that anticipated at steady state conditions and it was used to fill the rest of the process vessels. The process liquor loop was closed on August 23, 1980 with the overflowing of the clarified liquor from the thickener into the thickener hold tank. On August 24, 1980, a leak at the bottom of the thickener tank forced the first major outage of the system. Penetrations near the support studs for the bottom plate of the thickener had occurred due to inadequate grouting.

Lining separation had also taken place at various locations in the tank. Repairs were completed on September 15, 1980. The system was restarted the same day following an outage of 529 hours (22 days). Subsequently, two outages occurred in the month of September, totalling 121 hours (5 days). They were due to the need to increase the active sodium concentration in the system as it was resulting in the generation of solids with poor settling characteristics. Limitations in the soda ash feedrate mechanism and problems with control valves maintaining low flowrates resulted in the system being filled with a dilute sodium solution. (High soda ash feedrates and low flows were necessary while filling the empty system during the initial startup and following the repairs of the thickener.)

On September 29, 1980, the system was shut down for another long outage totalling 367 hours (15.3 days). The initial cause for the shutdown was the excessive vibration in the F.D. fan, but other mechanical problems extended the outage. Mechanical repairs included fixing fan bearings and balancing the fan; patching a hole in the scrubber and relining the affected area; repairing the thickener rake lifting mechanism after a failure of the limit switch caused the shaft to be driven through the housing; and resurfacing the leaky bottom of the first reactor. Repair work was completed by October 11, 1980; the system, however, could not be restarted until October 14, 1980 due to boiler water preheater problems in Gulf Power's boiler.

On October 20, 1980, CEA filed a petition for financial reorganization pursuant to Chapter 11 of the U.S. Bankruptcy Code. The system was shut down on October 24, 1980 to fix a broken end of the filter drum, as solids accumulated in the thickener were deteriorating and their filtering was imperative. Before repairs to the filter were completed, the impact of CEA's bankruptcy declaration was felt as the various organizations and subcontractors involved had to resolve legal issues before resuming normal operation.

The legal issues were resolved by November 24, 1980. Various mechanical repairs needed were performed then, and the system was restarted on November 28, 1980 following an outage of 853 hours (35.5 days).

In the month of December, the system was finally operated under stable conditions, processing flue gas for 565 hours (23.5 days). SO_2 removal efficiencies could not be monitored as the SO_2 analyzers were not yet in working condition, and the O_2 analyzers had not been installed. Occasional measurements with Drager tubes indicated SO_2 removal efficiencies in excess of 90%. The most significant causes of downtime in December were the difficulties in procuring raw materials. Two outages, due to lack of soda ash, amounted to 173 hours (7.2 days) of downtime.

In the month of January 1981, the system processed flue gas for 323 hours (13.5 days). Most of the downtime in this month were due to problems caused by solids carryover in the thickener overflow. The poor settling behavior of the waste solids, formed in the reactors during this period, was enhanced by piping and flow problems in the thickener. The lines leading to the thickener centerwell were found to extend as far down as the centerwell itself. Thus, these lines were actually stirring the bottom of the thickener; and the centerwell was, for all practical purposes, nonfunctional. Furthermore, a significant amount of channeling was observed to be taking place in the thickener; the liquor was effectively overflowing on only one half of the weir perimeter.

These problems were corrected by February 1, 1981; and with the SO_2 monitors becoming operational on January 18, 1981, the startup and break-in testing period was concluded at the end of January 1981.

B. SYSTEM TESTING

Having completed the piping changes and thickener overflow weir leveling mentioned above, the system began processing flue gas on February 2, 1981. The testing period extended until March 28, 1981.

During the testing period, the system operated for 925 hours (38.5 days) or 71.4% of the time. Significant outages were due to bad weather, filter repairs, and solids carryover in the thickener overflow. Heavy rains washed out the roads to the waste disposal area; and thus made it impossible to truck more waste to the ponds. The area underneath the filter and pug mill was already filled with waste material. This outage took place on February 7, 1981 and lasted 39 hours (1.6 days).

On March 5, 1981, the system was shut down in order to repair the filter. The performance of the vacuum filter was deteriorating steadily. Severe leaks had been detected in the internal piping of the filter drum; and by the time the system was shut down, only half of the panels were able to pull cake. This outage lasted 190 hours (7.9 days).

Following the repairs to the filter, significant problems with resuspension of the solids left in the thickener were encountered. The end result was the carrying of fines solids in the thickener overflow. These fines, if uncontrolled, can contribute to the formation of poor settling solids in the reactor. This subject is addressed in the Process Performance Section. The excessive carryover of solids caused outages totalling 94 hours (3.9 days).

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During the testing period, the system averaged an SO_2 removal efficiency of 95.8%; and it recorded an uninterrupted period of operation of 431 hours (18 days).

V. SYSTEM PERFORMANCE

The primary purpose of this program was to evaluate the performance at a prototype scale of the limestone dual alkali technology in order to determine its technical feasibility and to supply technical information related to the implementation of the process at a full, commercial scale. As such, the following discussion emphasizes the process performance since it is the single most important factor in determining the technical feasibility of the technology. System utilization and mechanical (equipment) performance are discussed in so far as they affected the process performance of the prototype system at Scholz.

A. SYSTEM UTILIZATION

The utilization of the system, as defined in this report, represents the number of hours the system operated and processed flue gas in relation to the total hours of the period under consideration. The utilization of the limestone dual alkali system throughout the program is graphically displayed in Figure V-1. It covers the overall operation of the system: startup and break-in period, and the system testing period. Details on system outages, their reason and duration are given in Appendix A.

The utilization of the dual alkali system was rather poor during the startup and break-in period; the system operated only 35% of the time. By the very nature of the startup and break-in operations, one would not expect a very high utilization during such period. At Scholz, however, additional complications contributed to the poor utilization of the system during this initial period. As some of these causes are not necessarily related to the technology, they must be kept in proper perspective.

10 Lines to End of Typing Area

As indicated in the discussion of the operating history of the system, 43% of the downtime during the startup and break-in period was attributable to mechanical problems, 20% to process problems, 26% to CEA's bankruptcy declaration, 7% to lack of raw materials, and 4% to miscellaneous causes.

In evaluating the utilization of the system, one must also consider the following facts. The only spare equipment at Scholz were pumps. A full-scale system would be modular in nature and would typically incorporate a spare module. Such redundancy would greatly facilitate the maintenance and repair of equipment without causing the loss of utilization of the overall system as was the case at Scholz. The utilization of the system was also affected significantly by the physical condition of the prototype unit. At the completion of the lime dual alkali test program in July 1976, the system was "mothballed" until recommissioning work began in 1979. The age of the equipment, the three years of inactivity, and the limitations of a finite recommissioning effort showed their effects during the operation of the plant making necessary an extensive maintenance effort.

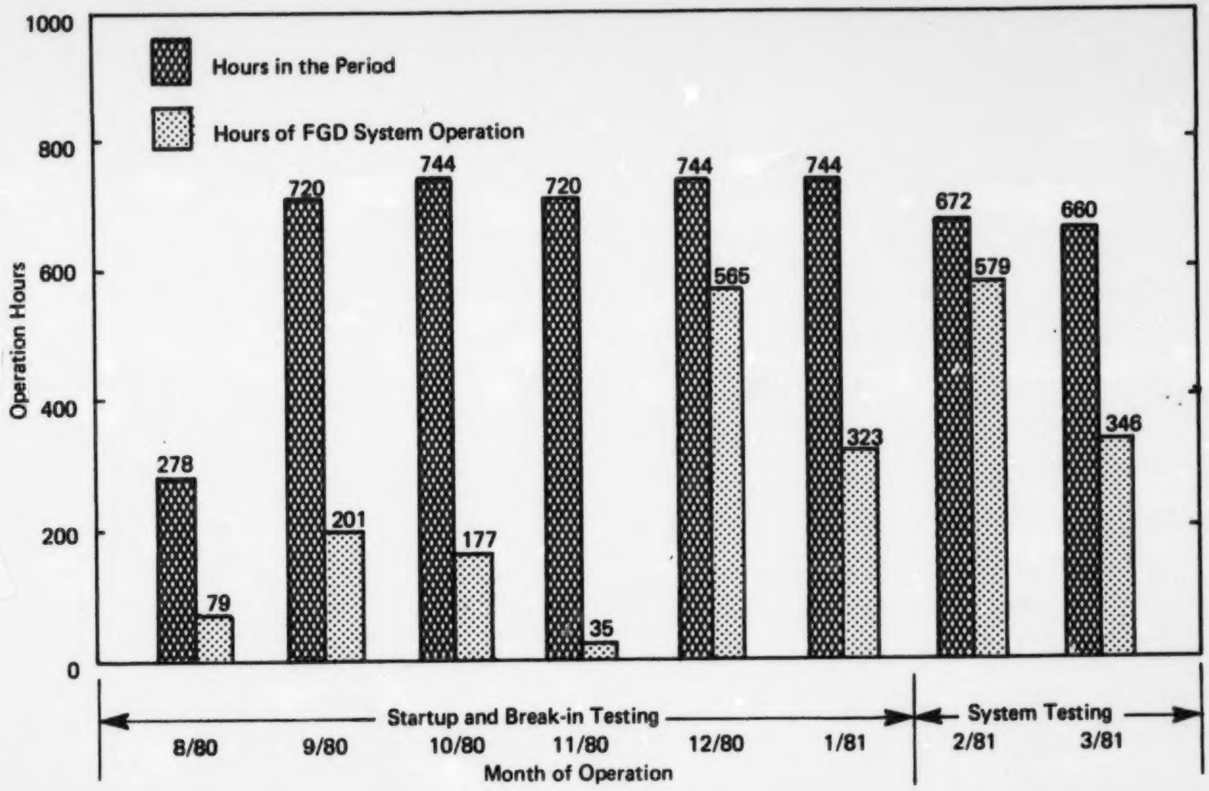

FIGURE V–1 AVAILABILITY OF THE LIMESTONE DUAL ALKALI SYSTEM AT SCHOLZ
(August 20, 1980 to March 28, 1981)

The losses of utilization due to the declaration of bankruptcy and to the lack of raw materials (due to reluctance of suppliers to provide materials in light of the declaration of bankruptcy) were only circumstantial and have no bearing on the technical merits of the process.

The utilization of the FGD system during the system testing period improved significantly. The system operated 71% of the time as the effect of some of the above mentioned extraneous factors decreased. Mechanical- and process-related problems accounted for most of the dowtime during the testing period; 63% due to mechanical problems, 30% due to process-related problems, and 7% due to miscellaneous causes.

B. PROCESS PERFORMANCE

The original intent of the test program was to evaluate the following aspects of the limestone dual alkali process performance:

- SO_2 removal capabilities;
- Consumption of raw materials;
- Consumption of energy;
- Levels of sulfite oxidation experienced by the system as well as the ability of the system to tolerate such levels of oxidation;
- Waste cake properties;
- Particulate matter removal (during periods of planned electrostatic precipitators turn down) by the system and the effect of ash on process chemistry;
- Sulfate emissions in the processed flue gas; and
- Reliability and ease of operation.

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to End of
Typing Area

The test program was to be conducted in two phases--a two month baseline testing and a four month variational testing. In addition, preliminary tests were to be performed during the initial startup and break-in phase which was anticipated to last two months. The information obtained in the preliminary testing would be used in optimizing and finalizing control set points and in verifying the suitability of the testing and analytical procedures.

Previous pilot plant and laboratory studies determined the general effect of process parameters and operating conditions on the performance of limestone dual alkali systems. The baseline testing of this program was to confirm and further characterize these relationships. Most importantly it was to provide a quantitative evaluation of the overall system while it was operated under a constant set of operating conditions and process parameters. Subsequent variations in selected process parameters and operating conditions and their effect on the performance of the system were to be studied in the variational testing phase. The results of these variational tests would provide the technical information needed to determine the range of application and limitations of the limestone dual alkali technology. As such, these tests were to include the type of variations expected in full-scale applications: boiler load, inlet SO_2 concentrations, type of limestone, boiler excess air, and particulate loading.

For reasons that have been previously discussed, the testing of the system was reduced to only two months--February and March 1981. Given the limited testing performed, a thorough and complete evaluation of the technology was not possible. Therefore, care should be exercised in drawing final conclusions from the results of this test program.

In order to provide even a small amount of additional information, the following discussion on the performance of the process also incorporates the information obtained during the break-in testing in December 1980 and January 1981.

1. Operating Conditions

The general operating conditions during the break-in period and the system testing period are summarized in Table V-1. Time plots detailing the operating conditions of the system are given in Appendix C. Process flows and stream compositions for selected days during the test period are given in Appendix D.

During the operation of the system, the boiler was fired with coal containing 2.6 to 4.2% sulfur (See Appendix B for additional coal characteristics). The resulting SO_2 concentrations in the flue gas typically ranged from 1400 to 2200 ppm. Injection of SO_2 prior to the gas entering the scrubber expanded this range to concentrations as high as 3240 ppm, although SO_2 concentrations usually did not exceed 2500 ppm. The gas load to the system varied between 13,300 and 48,800 dry scfm (equivalent to 6 to 23 MW). Oxygen levels in the flue gas varied between 5 and 11.2 vol.% depending on boiler load.

10 Lines to End of Typing Area

The liquor in the system had an active sodium concentration[1] normally ranging between 1.4 and 1.7 M, occasionally reaching extremes of 1.1 and 1.8 M. The sulfate concentration reached a steady level of about 1 M. Similarly, the chloride concentration leveled out at 0.05 to 0.07 M.

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[1] The term "active sodium" (Na^+ active) is equivalent to the concentration of the following species: $[NaHSO_3] + 2x[Na_2SO_3] + 2x[Na_2CO_3] + [NaHCO_3]$. As such, it is an indirect indication of the absorptive capacity of the liquor.

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TABLE V-1

SUMMARY OF SYSTEM OPERATING CONDITIONS

(December 1980 - March 1981)

Inlet Gas	Range	Average
Gas Load (dry scfm)	13,300 - 48,800	38,100
SO_2 Level (ppm, dry basis)[a]	1,460 - 3,240	2,071
O_2 Level (% dry vol.)[a]	5.0 - 11.2	7.1

Regenerated Liquor Composition	Range	Typical
pH	5.70 - 6.40	6.10
Na^{+}_{act} (M)	1.1 - 1.8	1.6
$SO_4^{=}$ (M)	0.76 - 1.07	0.95
Cl^{-} (M)	0.030 - 0.087	0.070
Mg^{++} (ppm)	600 - 900	750
Ca^{++} (ppm)	14.5 - 37.5	22.0

---

[a] Based on data from February and March only, since SO_2 and O_2 meters were unavailable during the other months.

The overall performance of the system in terms of SO_2 removal, chemical requirements, and properties of the waste cake is summarized in Table V-2. The performance data is based on overall material balances derived from flue gas analyses, waste cake properties and discharge rates, and raw materials feedrates and inventories.

2. SO_2 Removal

The SO_2 removal performance of the system was excellent. Removal efficiencies were monitored continuously during the testing period--February and March 1981; only partial information was collected in January as final adjustments and calibration of the SO_2 monitors were completed in mid-January. A summary of the SO_2 removal performance of the system is given in Table V-3.

Inlet SO_2 concentrations during the testing period ranged from 1460 to 3240 ppm (dry volume basis). Outlet SO_2 concentrations, corrected for reheater air dilution, ranged from 29 to 238 ppm. The corresponding SO_2 removal efficiencies averaged to 95.4% for the month of February and 96.7% for the month of March. During the break-in period, occasional measurements were made in the month of December 1980 using Drager tubes. These measurements indicated SO_2 removal efficiencies in excess of 90%. Removal efficiencies during the time the monitors were in operation in January averaged to 93.5%.

The single most important variable affecting the SO_2 removal was the liquor pH. This strong dependancy is clearly shown in Figure V-2 and V-3. Outlet SO_2 concentrations less than 100 ppm were obtained at scrubber bleed pH's higher than 5.7; concentrations less than 50 ppm were obtained at a pH of 6. Similarly, SO_2 removal efficiencies greater than 90% could easily be achieved by maintaining a scrubber bleed pH of at least 5.5. Although the SO_2 removal capability improves as the scrubber bleed liquor pH is raised, it begins to level off at a pH of 6. Furthermore, regeneration reactions are carried out at pH's of 6.0 to 6.4. Thus, a limitation is needed on the scrubber bleed pH in order to maintain an adequate control on these reactions and to ensure sufficient reactivity of the liquor. Operating the scrubber/absorber at a bleed pH of 5.7 to 6.0 appears to be a reasonable compromise that provides for high SO_2 removal efficiencies and adequate regeneration of the liquor in the reactors.

The scrubber bleed pH is strongly dependent on the ratio of $[HSO_3^-]/[HSO_3^- + SO_3^=]$ concentrations in the liquor; this dependancy is shown in Figure V-4. Thus, one could also use these concentrations as a guideline for SO_2 removal. During the test period, the ratio of $[HSO_3^-]/[HSO_3^- + SO_3^=]$ typically ranged between 0.60 and 0.75, yielding SO_2 removal efficiencies between 94 and 98%.

The impact of other variables such as inlet SO_2, sodium concentration, and pressure drops on the SO_2 removal performance is much smaller than the bleed pH. Their effect falls within the scatter of the data and as such appear to be secondary in nature.

TABLE V-2

SUMMARY OF OVERALL SYSTEM PERFORMANCE

(December 1980 - March 1981)

Month	Coal Sulfur Content (wt.%)		SO_2 Removal Efficiency	Limestone Feed Stoichiometry (mols $CaCO_3$/ mol ΔSO_2)	Soda Ash Feed Stoichiometry (mols Na_2CO_3/ mol ΔSO_2)	Waste Cake % Solids (wt %)	
	Range	Average				Range	Average
Break-in Period							
December	2.8-3.6	3.3	--[a]	--[a]	--[a]	33.6-47.7	37.0
January	3.0-3.9	3.3	93.5%[a]	0.83[a]	0.20[a]	36.3-52.0	42.0
System Testing							
February	2.6-4.2	3.5	95.4%	0.82	0.26	33.3-46.6	38.0
March	3.3-3.9	3.6	96.7%	0.67	0.33	32.2-42.7	38.0

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[a] SO_2 analyzer inoperable until mid-January.

TABLE V-3

SO_2 REMOVAL PERFORMANCE

Absorber and Scrubber Conditions	Break-in Testing		System Testing	
	December 1980	January 1981	February 1981	March 1981
Scrubber ΔP (in. H_2O)	6.1 - 7.9	5.4 - 8.6	5.8 - 10.1	7.2 - 10.3
Absorber Tray ΔP (in. H_2O)	1.3 - 4.0	1.2 - 4.8	1.3 - 3.2	1.3 - 2.5
Scrubber Inlet SO_2 (ppm)[b]				
Range	1,400 - 2,150[a]	1,400 - 3,200	1,850 - 2,640	1,460 - 3,240
Average	1,650[a]	1,862	2,036	2,142
Absorber Outlet SO_2 (ppm)[b]				
Range	---[d]	42 - 334	38 - 238	29 - 116
Average	---[d]	122	94	71
Average Scrubber SO_2 Removal Efficiency (%)[c]	---	---	65.8	69.4
Average Absorber SO_2 Removal Efficiency (%)[c]	---	---	29.6	27.3
Average Total SO_2 Removal Efficiency (%)	---[d]	93.5	95.4	96.7

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[a] Based on Gulf Power's SO_2 analyzer.
[b] By volume, dry basis.
[c] Average removal efficiencies obtained from laboratory analysis.
[d] SO_2 analyzer inoperable.

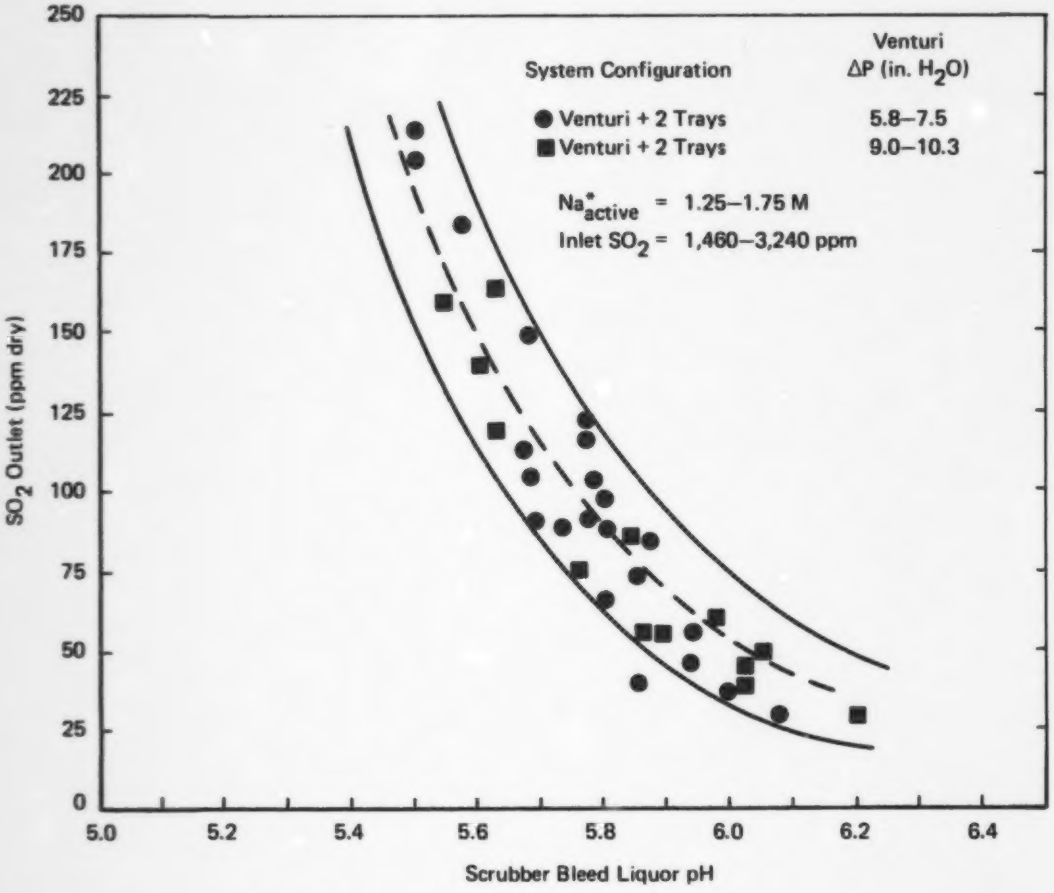

FIGURE V-2: OUTLET SO$_2$ AS A FUNCTION OF SCRUBBER BLEED pH

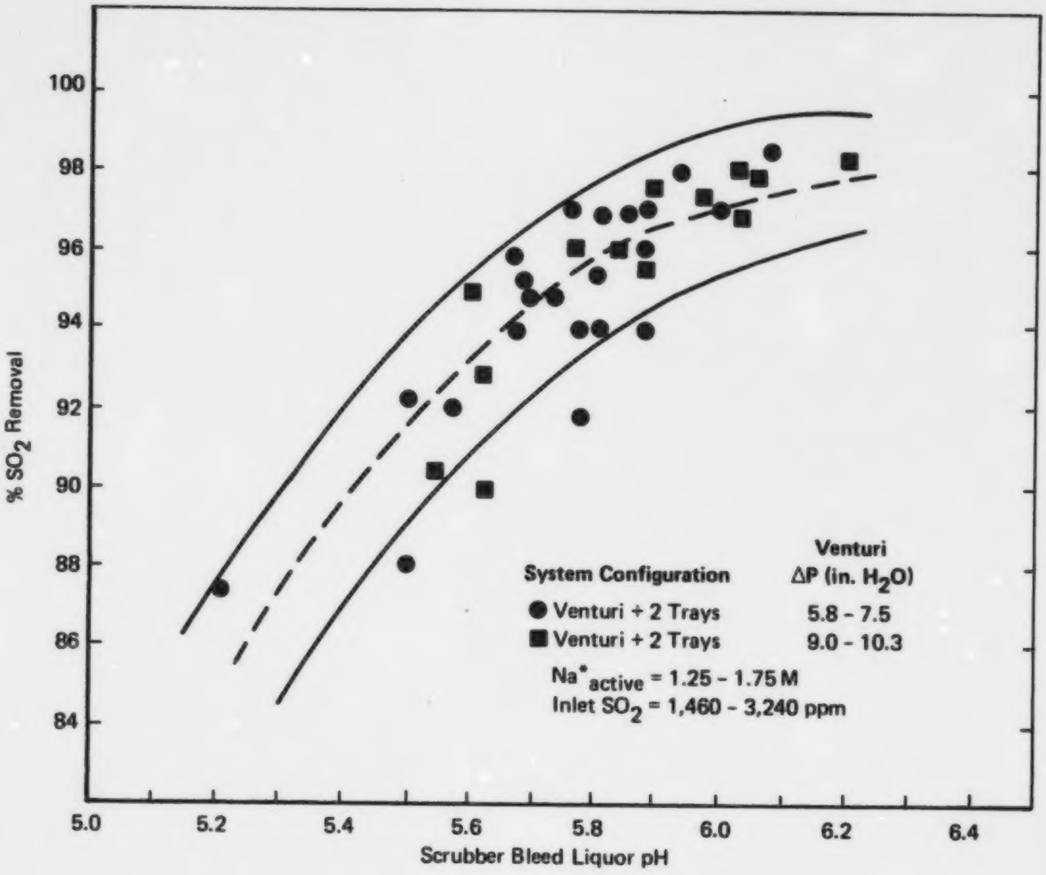

FIGURE V–3 SO_2 REMOVAL AS A FUNCTION OF SCRUBBER BLEED pH

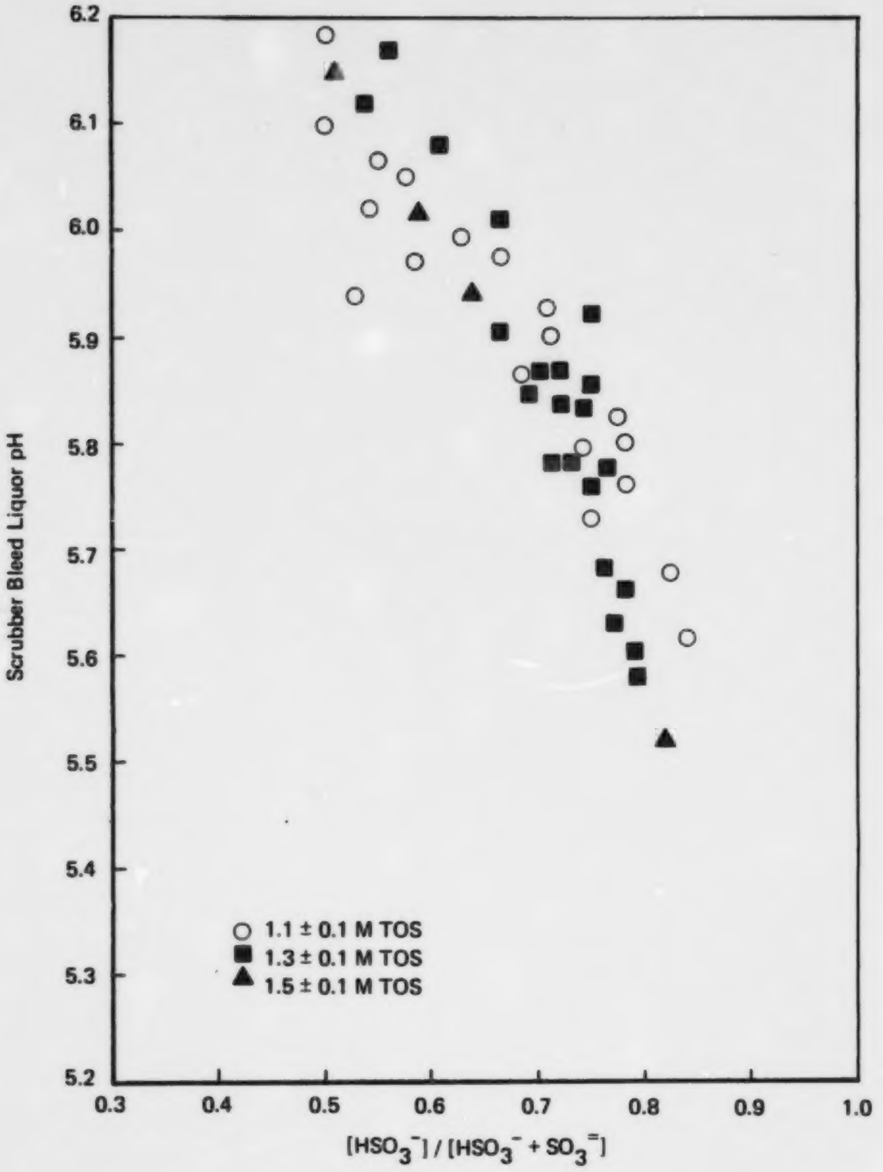

FIGURE V-4: pH VS. SOLUTION COMPOSITION – SCHOLZ PLANT (SCRUBBER EFFLUENT)

3. Limestone Utilization

The limestone utilization by the dual alkali system was very good. During the testing period, utilizations in the reactor train effluent ranged from 85 to 95% of the available $CaCO_3$ in the raw limestone. As the reaction with limestone continued in the dewatering system, the final system utilizations, determined from chemical analyses of the filter cake, ranged from 93 to 100%, averaging at 97.5%.

Two types of limestone were used for regenerating the spent scrubber liquor. Fredonia limestone from Kentucky was used during the startup and break-in period until the end of December 1980. Thereafter, Sylacauga limestone from Alabama was used for the remainder of the startup and break-in period and for the entire testing period. The characteristics of these limestones are given in Table V-4. The Fredonia limestone was selected for use in the initial period of operation of the system since the successful results in the previous laboratory and pilot plant test programs[2] were obtained with this limestone. Following the stable operation in the month of December, a new limestone was used to test the capability of the system with regard to the use of different limestones. The information below presents the results obtained during the testing period in which Sylacauga limestone was used.

The limestone utilization and feed stoichiometry during the testing period is summarized in Table V-5. Overall system utilization of limestone was determined by two independent methods--from the filter cake analysis and from material balances based on limestone feed stoichiometries, soda ash consumption and SO_2 removed. The corresponding utilizations for the testing period were 97.5% and 94.2%, respectively. The agreement between these results is quite reasonable considering the normal limitations of the analytical procedures and instrumentation used.

Limestone feed stoichiometries (mole of $CaCO_3$ fed per mole SO_2 removed) were determined by using information from the feedrate control instrumentation. The average limestone feed stoichiometry for the testing period was 0.77 moles of $CaCO_3$ per mole of SO_2 removed. Two independent checks of this result were made. The first using the overall limestone consumption based on truck deliveries and silo inventories; and the second based on chemical analysis of the filter cake. The corresponding stoichiometry values of 0.68 and 0.70 provide a reasonable check.

10 Lines to End of Typing Area

Analyses were also made to determine the progress of the regeneration reactions, and thus limestone utilization, through the reactor train. The results of these analyses are shown in Figure V-5. The effect that the solids carried over in the thickener overflow liquor (and thus fed to the reactor train after passing though the absorber and scrubber) had on the limestone utilization in the first reactor is of particular interest. The utilization in this reactor ranged from 23% at solids

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[2] LaMantia, C.R., et al., op. cit., Volume II.

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TABLE V-4

LIMESTONE PROPERTIES

	Fredonia	Sylacauga
Supplier	Fredonia Quarries	Sylacauga Calcium Products
Mine Location	Fredonia, Kentucky	Sylacauga, Alabama
Type	High-Calcium Calcite	High-Calcium Calcite
Period of Use	Aug. 80 - Dec. 80	Jan. 81 - Mar. 81
Particle Size Distribution Wet Sieve Analysis (wt.%):		
<60 mesh (250μ)	99.7	99.9
<80 mesh (177μ)	99.5	99.8
<100 mesh (149μ)	99.4	99.8
<150 mesh (104μ)	99.2	99.8
<230 mesh (61μ)	98.4	99.6
<325 mesh (44μ)	96.4	97.7
<400 mesh (38μ)	94.4	93.9
Bulk Density (lbs/ft^3)	55-60	60-65
Chemical Analysis		
Alkalinity (wt.% as $CaCO_3$)	97.9	97.8
Ca (wt.% as $CaCO_3$)	93.6	95.7
Mg (wt.% as $MgCO_3$)	4.2	1.3
Fe (wt.% as Fe_2O_3)	0.05	0.04

TABLE V-5

LIMESTONE UTILIZATION PERFORMANCE

Period	Hours of Operation	Limestone Utilization (Percent of $CaCO_3$)			Limestone Feed Stoichiometry ($CaCO_3/\Delta SO_2$)[c]		
		Reactor T-5[a]	Filter Cake[a]	From Limestone Feed Stoichiometry[b]	Feed Control Instrumentation	Deliveries and Inventory	Filter Cake
January 28 - March 3	644	86.0	97.0	94.0	0.79	0.66	0.72
March 4 - March 24	197	92.0	98.0	96.0	0.65	0.82	0.63
January 28 - March 24	841	89.5	97.5	94.2	0.77	0.68	0.70

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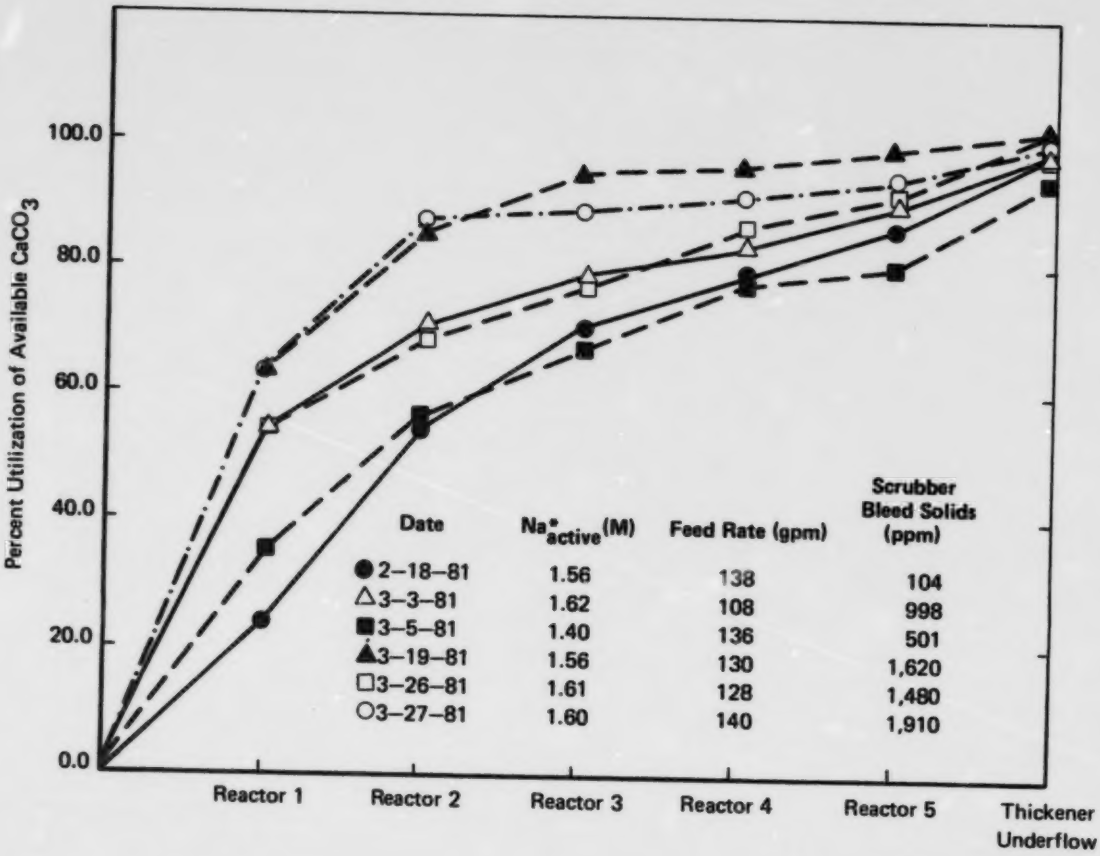
[a] These figures represent the average of the number of samples analyzed in the laboratory.

[b] Based on feed control instrumentation record.

[c] Stoichiometry averages for the overall period are weighed averages based on the amount of SO_2 removed per individual period.



Date	Na^*_{active} (M)	Feed Rate (gpm)	Scrubber Bleed Solids (ppm)
● 2–18–81	1.56	138	104
△ 3–3–81	1.62	108	998
■ 3–5–81	1.40	136	501
▲ 3–19–81	1.56	130	1,620
□ 3–26–81	1.61	128	1,480
○ 3–27–81	1.60	140	1,910

FIGURE V-5: LIMESTONE UTILIZATION FOR REACTOR STUDIES

carryover of 104 ppm to 63% at 1910 ppm. Since the solids carried over in the thickener overflow were essentially fully reacted, they undoubtedly contributed to an "apparent" high limestone utilization in the first reactor. In addition, it is also possible that solids entering the first reactor acted as seed crystals facilitating and increasing the rate of precipitation of calcium sulfur salts. This, in turn, would have promoted an increase in the rate at which calcium from the limestone dissolved in the liquor and reacted with the sodium bisulfite. Further studies would be needed to verify this hypothesis. As reacting slurry passed through the reactors, the holdup time in the first reactor became small ($\sim$5 min.) in relation to the overall holdup time in the reactor train ($\sim$100 min.), and the spread reduced to limestone utilizations of 82 to 99% in the last reactor. Additional reaction in the thickener further reduced the differences in limestone utilizations.

Thus, it seems that regardless of the extent of reaction achieved in the first reactor, the limestone will be efficiently consumed in the remainder of the reactor train and dewatering system. The solids carryover in the thickener overflow do not appear to have a substantial effect on the overall limestone utilization by the system; they do, however, impact on the settling properties of the solids generated as will be discussed later when addressing the properties of the waste solids.

Further characterization of the reactor system and limestone behavior is necessary as a result of the limited testing performed in this program. As previously indicated while describing the system at Scholz, single reactor configuration was used in the test period--five reactors in series with dry limestone feed to the first reactor. Although a second type of limestone--Fredonia--was used during the startup and break-in period, the information available was not sufficient to characterize its use. Only a limited amount of testing was being performed during that period and SO_2 instrumentation was not available. Nevertheless, based on the stable operation achieved in December 1980, using the Fredonia limestone, all indications were that this limestone exhibited performance characteristics comparable to the Sylacauga limestone used during the testing period.

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Typing Area

4. Oxidation and Sulfate Precipitation

Oxidizable sulfur species in the process liquor (namely, sodium sulfite and bisulfite) can be oxidized to sodium sulfate whenever exposed to oxygen. Given the need for intimate contact between the liquor and the flue gas in order to effect adequate SO_2 removal, one would anticipate that oxidation would take place primarily in the SO_2 absorption units and to a lesser extent throughout the rest of the system. Oxidation during SO_2 absorption would, of course, depend on the oxygen content of the flue gas, temperature, and gas/liquid contact characteristics.

The overall level of oxidation in the system is measured by the amount of SO_2 removed from the flue gas which leaves the system in the oxidized state (sulfate). At Scholz, the levels of oxidation experieced in the absorber/scrubber unit typically ranged from an equivalent of 10 mole %

of the SO_2 removed, at oxygen concentrations of 6 vol.% in the inlet flue gas, to 25 mole % at oxygen concentrations of 8%. These results are shown in Figure V-6. The scatter of data, within a band equivalent to ± 5 of ΔSO_2, represent variations in temperature, pressure drop, SO_2 concentration in the gas, concentration of oxidizable sulfur species in the liquor, as well as the limitations in accuracy of sampling and analytical procedures. More extensive data would be required to differentiate among these effects.

Oxidation throughout the remainder of the system amounted to an additional 1 to 5% of the SO_2 removed. Further oxidation may take place while the solids are being removed from the system by the filter. Such oxidation is of secondary importance as it does not affect the operation of the system.

During the stable operation of the system, the sulfate formed in the scrubber/absorber unit, as well as the rest of the system, must be removed at the same rate at which it is generated. This removal is effected by the precipitation of calcium sulfate and its discharge as part of the cake solids and by the losses of sodium sulfate in the entrained liquor in the cake.

Although the process liquor is unsaturated with respect to calcium sulfate, this component is co-precipitated along with the calcium sulfite. The extent of this co-precipitation is dependent on the relative concentrations of sulfate and sulfite ions in the liquor, as shown in Figure V-7. It is interesting to note that the limestone-based dual alkali seems to be better able to co-precipitate sulfate than the lime-based dual alkali system. Co-precipitation in the limestone-based system may, however, be limited by the need to maintain a high active sodium concentration which will limit sulfate concentrations (and consequently the sulfate/sulfite ratio) simply by solution saturation consideration. Furthermore, the sulfate/sulfite ratio may also be limited by the need to ensure high limestone utilizations and good solids properties. The sulfate/sulfite ratio in the regenerated liquor at Scholz did not exceed a value of 4. Additional testing would be required to determine a practical limit for this ratio and the corresponding limitation in the ability of the system to co-precipitate sulfate.

10 Lines
to End of
Typing Area

Sulfate balances, which are summarized in Table V-6, were performed over two separate periods. The calculated sulfate formation in the system is based on the oxidation estimated to have taken place in the absorber/scrubber section as well as in the reactor and dewatering sections. The fate of the sulfate formed was independently accounted for by determining the amount of sulfate discharged in the cake, the change in sodium sulfate inventory in the system, and the miscellaneous liquor losses from the system.

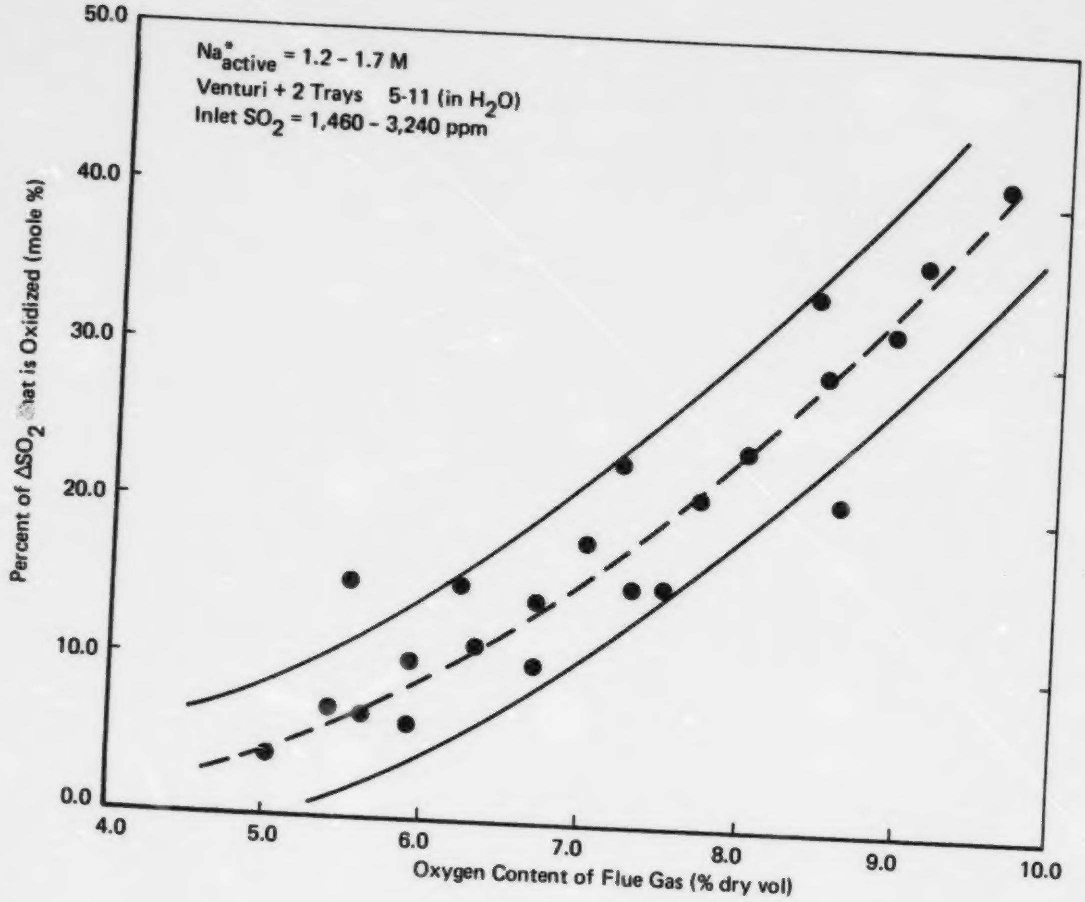

FIGURE V–6 OXIDATION IN THE SCRUBBER/ABSORBER SECTION AS A FUNCTION OF THE OXYGEN CONTENT IN THE INLET FLUE GAS

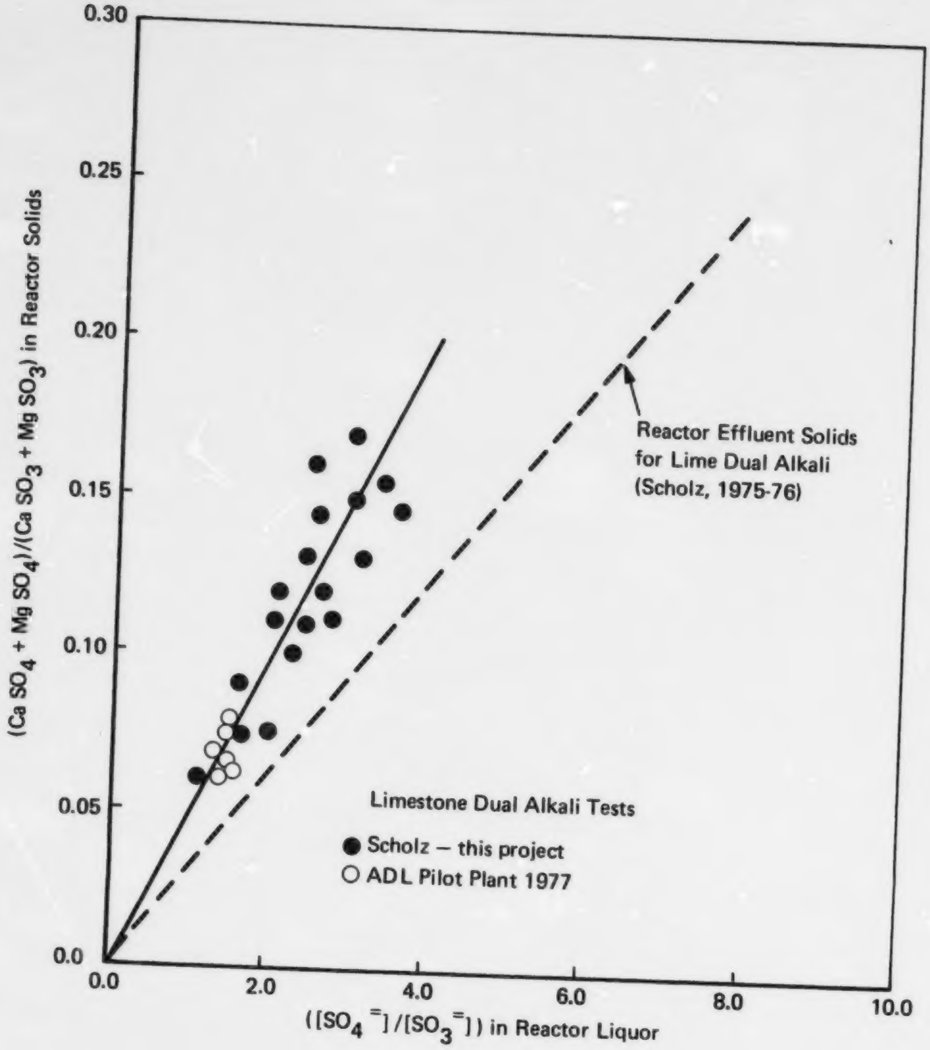

FIGURE V-7 : $(Ca\ SO_4 + Mg\ SO_4)/(Ca\ SO_3 + Mg\ SO_3)$ RATIO IN REACTOR EFFLUENT SOLIDS AS A FUNCTION OF $(Na_2\ SO_4/Na_2\ SO_3)$ RATIO IN REGENERATED LIQUOR

TABLE V-6

SYSTEM SULFATE BALANCE

	Period	
	January 28 - March 4	March 4 - March 24
Hours of Operation	644	197
Average Inlet SO_2 (ppm)[a]	2,013	1,966
Average SO_2 Removal (%)	95.7	96.3
Average Inlet O_2 (vol.%)	7.9	6.7
Sulfate Formation (% ΔSO_2)	18.0	17.0
Sulfate Account (% ΔSO_2)		
In Cake - $CaSO_4$	6.3	6.5
Na_2SO_4	3.4	2.9
ΔNa_2SO_4 in Liquor	2.5	-4.4
Liquor Losses[b]	4.8	11.1
Total Accounted for (% ΔSO_2)	17.0	16.1

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[a] By volume, dry basis.

[b] Includes sulfate losses from pump seal leaks, inadvertent spills, and any purges to the pond. It is calculated independently of the total sulfate balance as follows:

Other = (soda ash makeup - sodium losses in the cake - accumulation of sodium in system liquor)

$$\times \left(\frac{\text{average solution sulfate concentration}}{\text{average total } Na^+ \text{ concentration}}\right)$$

During the first period (January 28, 1981 to March 4, 1981), the sulfate formation averaged at an equivalent of 18% of the SO_2 removed. Of the sulfate formed, the equivalent of 9.7% of the SO_2 removed left the system with the cake; 2.5% were accounted for in the change in composition of the system liquor; 4.8% were accounted for in liquor losses, and 1% is unaccounted for. Thus, over 94% of the estimated sulfate formed is accounted for, which provides a reasonable check of the estimated level of oxidation in the system. Similarly, during the second period (March 4, 1981 to March 24, 1981), the sulfate formation averaged at 17% of the SO_2 removed. The equivalent of 9.4% of the SO_2 removed left the system with the cake. The sodium sulfate inventory in the system liquor declined in an amount equivalent to 4.4%. Liquor losses accounted for the equivalent of 11.1% of the SO_2 removed, and an equivalent of 0.9% of the SO_2 removed was unaccounted for.

The liquor losses from the system were indeed severe. Numerous leaks were encountered throughout the entire operation: pumps, valves, corroded carbon steel piping for the soda ash solution lines, etc. The age and condition of the equipment, even after recommissioning, were primarily responsible for these leaks. The small capacity of the thickener hold tank (1500 gallons vs. 93,000 gallons in the thickener) and its lower overall elevation in relation to the thickener resulted in several inadvertent spillages. Excessive inputs of seal water (needed to maintain the old pumps in operation) combined with heavy rains and low evaporation rates in the scrubber (because of low inlet gas temperature due to low boiler loads and high boiler leakages) caused severe water balance problems. Purges of process liquor had to be made to Gulf Power's ash pond whenever the water balance in the system could not be maintained and spillages were imminent.

A properly designed full-scale unit would incorporate collection sumps to return liquor losses due to leaks and spills back to the system. It would also provide adequate surge capacity to handle water balance problems. In that case and for the same level of oxidation in the system, the sulfate concentration in the system would increase. The sulfate/sulfite ratio in the liquor would also increase correspondingly; and, as shown in Figure V-7, a higher precipitation of $CaSO_4$ would be effected thereby re-establishing the equilibrium between sulfate formation and its removal from the system.

10 Lines to End of Typing Area

5. Waste Solids

a. Settling Characteristics

The generation of solids with good settling characteristics was the most significant process limitation encountered at Scholz. It essentially accounted for all of the process related outages. Throughout the months of December and February, solids with excellent settling characteristics were generated. The solids settled out to 10% of the initial slurry volume in 6 to 8 minutes. In contrast, the poor solids generated throughout January took hours rather than minutes to settle. Variations in the settling characteristics of the solids in the reactor train effluent are shown in Figure V-8. A definite correlation exists between the settling behavior of the reactor train effluent solids and the amount of solids in the scrubber bleed fed to the reactor train. The carryover of fine solids in the thickener overflow, which were fed to the reactors after passing through the absorber/scrubber, promoted the formation of more fine and difficult to settling solids in the reactor train. This was verified in the laboratory by running parallel regeneration tests with a scrubber bleed sample containing fine carryover solids and a similar sample from which the solids had been removed in a centrifuge. The settling rate of the solids formed by regenerating the first sample was noticeably slower than the one for the second sample. Micrographs of solid samples taken from the reactor train effluent are shown in Figure V-9. It is clear, from the observation of these pictures, that the poor settling behavior observed on March 27, 1981 was due to the large percentage of fine, needle-shaped solids present. On the other hand, the large agglomerated particles formed on February 18, 1981 have a much lower surface area to mass ratio and thus exhibited a much higher settling rate.

In general, a solids carryover in the thickener overflow of 1000 ppm could easily be handled by the system without any detrimental effects to the settling properties of the reactor solids. Solids carryover in excess of 5000 ppm, however, resulted in the quick deterioration of the quality of solids generated in the reactors.

Another characteristic of the waste solids was the deterioration of their settling properties with time. This effect was observed on various occasions, most notably during the filter repair downtime in March 1981. Solids with excellent settling properties had been generated in February and in the first few days of March. The system was shut down for filter repairs on March 5, 1981. Solids had accumulated in the thickener precisely because of the inability of the filter to effectively remove solids. During the eight days of downtime, the solids in the thickener began resuspending as the underflow was continuously recycled to the thickener centerwell to prevent plugging of the lines or the thickener discharge cone. It is possible that these agglomerated and quick settling solids began redissolving and recrystallizing, this time into needle-like, poor settling solids.

10 Lines to End of Typing Area

Occasionally, a flocculant[3] had to be used in order to aid in the settling of poor solids. Nevertheless, the use of flocculants as normal additives to the system does not seem warranted given the stable operation that was achieved throughout December and February without their use. It is sensible, however, to incorporate, in any future application, a simple system for addition of flocculant during periods of unforeseen generation of poor solids.

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[3] Hercofloc 1031 manufactured by Hercules, Inc., Wilmington, Deleware.

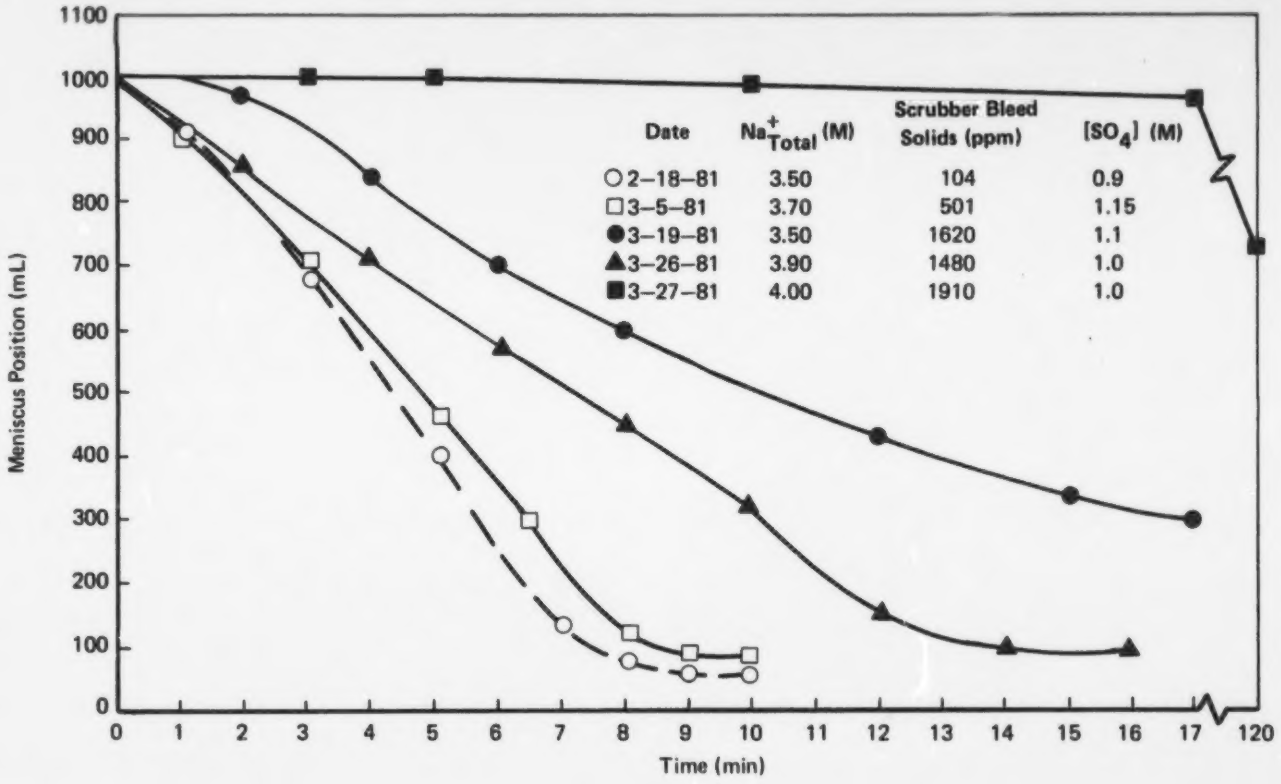
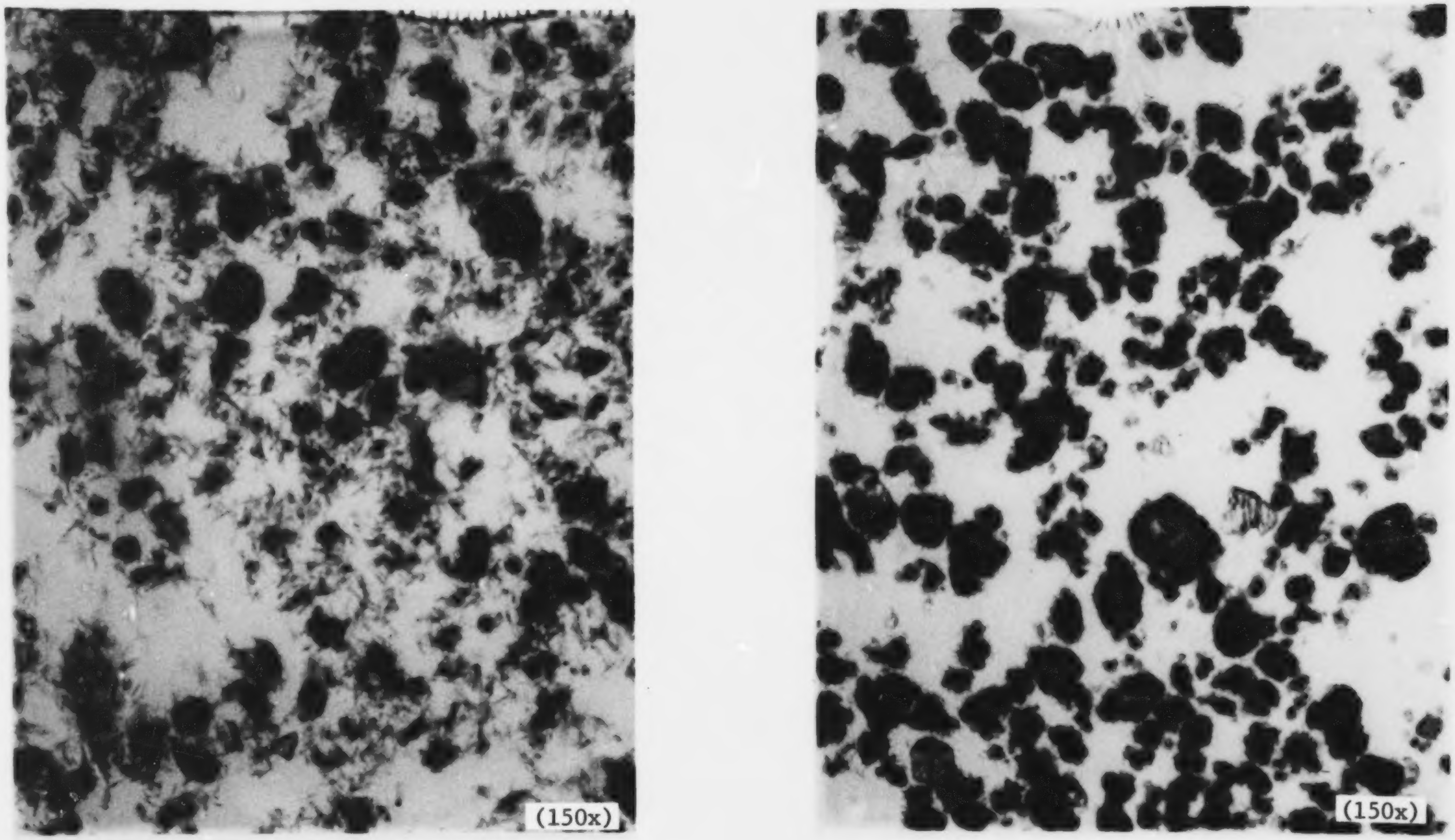


FIGURE V–8 VARIATIONS IN SETTLING RATES OF REACTOR T–5 EFFLUENT SOLIDS



Poor Settling Characteristics 3-27-81

Excellent Settling Characteristics 2-18-81

Figure V-9: Scanning Electron Micrographs of Reactor Train Effluent Solids

b. Filter Cake Characteristics

Ease of handling and disposing of the waste as well as economic considerations dictate the need for a reasonably dry waste material from which most of the sodium value has been recovered. In general, both of these goals were not achieved at Scholz. This does not imply that the problem is inherent to the limestone dual alkali technology as many mechanical and operational problems contributed to this shortcoming.

The anticipated insoluble solids content in the cake was 55 wt.% or higher. Yet, throughout the test program, it typically ranged between 35 to 45%. On a few occasions, it reached as high a solids content as 52%, and as low as 32%. The poor mechanical performance of the filter contributed significantly to this problem. Leaks in the internal piping of the filter drum and in the automatic control valve caused substantial losses of vacuum. The properties of the cake itself, which tended to crack during the dewatering cycle (see Figure V-10) also contributed to the loss of vacuum. (The use of standard devices such as rollers or flappers may alleviate this problem.) The end result was vacuum levels which typically ranged from 5 to 10 in. Hg instead of the 15 to 20 in. Hg that had been anticipated.

Another factor was the inability of the thickener underflow pumps to handle slurries containing much more than 15% solids. This required the dilution of the thickener underflow before being pumped to the filter. A much higher demand was then placed on the already limited capabilities of the filter. The slurry had to be dewatered from 15% solids to 55% instead of from 25 to 55%. The volume of filtrate to be handled more than doubled while removing the same amount of insoluble solids. Unfortunately, the filtrate pump and associated piping was not designed to remove this high amount of filtrate. This problem was further complicated by the internal leaks in the filter drum which caused the filtrate to be squirted back into the filter cake during the blowback cycle in which compressed air is used to release the cake from the drum surface.

10 Lines to End of Typing Area

Considering the above facts, it seems that with a properly working and adequately sized filter and associated pumps and piping, the waste cake would indeed have had a better chance of reaching the target 55% solids concentration, or even exceed it as was the case in the previous pilot plant tests.

The sodium losses in the cake were rather high throughout the test program. The loss of soluble sodium salts in the cake is reduced by maximizing the solids content in the cake and by using wash water to displace the remaining mother liquor in the cake. At Scholz, two spray banks were used to wash the cake. The use of up to 40 gpm of wash water was anticipated (equivalent to displacing four times the final volume of liquid in cake containing 55% solids). The corresponding sodium losses were anticipated to amount to about 4 wt.% of the insoluble solids content of the cake (equivalent to a Na/Ca ratio of 0.08 in the final cake).

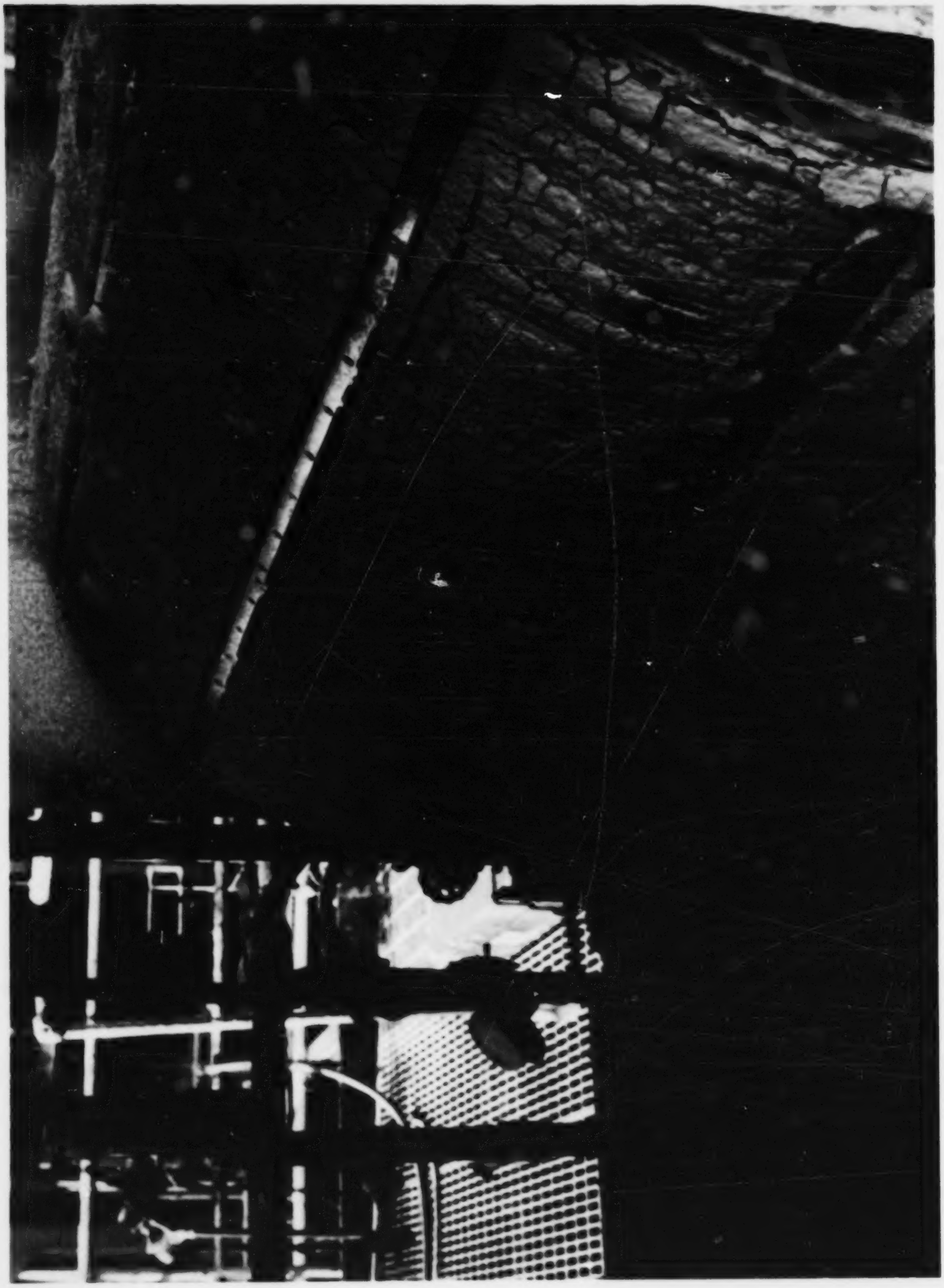


Figure V-10: Discharge of Vacuum Filter Cake

Filter cake properties are summarized in Table V-7. The average number of displacement washes in February was only 1.5 resulting in a high Na/Ca ratio of 0.2 in the discharged cake. In March, the average number of displacement washes was even lower, 0.8, and the Na/Ca ratio was correspondingly higher, 0.4. There were times, however, when up to three displacement washes could be accommodated; the sodium losses were then reduced to 0.03 to 0.04 moles Na/mole Ca, even below the targeted 0.08 moles Na/mole Ca. In general, the use of more than two displacement washes was not possible for three reasons: physical limitations in the handling of filtrate, limitations in the wash water supply, and the operation of the waste disposal system.

The limitations in filtrate handling capabilities have been discussed above. Obviously, further increasing the volume of filtrate by adding more wash water could only aggravate the problem.

Initially, well water was used for washing the cake, for pump seals as well as for instrumentation purges. The continuous deterioration of the pumps increased the demand for seal water, thus limiting the supply available for washing the cake. This deficiency was corrected during the testing period by changing the wash water supply from well to river water.

Finally, operational problems in the waste disposal system in the handling of the volume of cake generated indirectly affected the proper washing of the cake. In order to accommodate the operation of the waste disposal system, the rate of discharge of the cake had to be reduced considerably. The resulting thin cake demanded the use of low wash rates in order to prevent the cake from being washed off the filter drum. This problem became severe in March 1981 when the waste disposal system became operational. The average wash rate for this month was equivalent to only 0.8 displacement washes.

Again, it appears that the problem of high sodium losses in the cake is not inherent to the limestone dual alkali technology, but was rather caused by the mechanical condition of the equipment used and by operational problems specific to this program. Both of these conditions could be corrected in any future applications.

10 Lines to End of Typing Area

6. Soda Ash Consumption

The dual alkali technology incorporates the feeding of soda ash in order to make up for sodium losses in the system. The design of the Scholz system anticipated a makeup rate of 0.04 moles of Na_2CO_3/mole of SO_2 removed. The actual sodium losses were much higher and were due to not only excessive losses in the cake but also severe liquor losses due to leaks, spills, and liquor purges needed to maintain volume balances. During the testing period, the overall soda ash feed stoichiometry amounted to 0.29 mole of Na_2CO_3/mole of SO_2 removed. As much as 50% of this amount was needed to make up for cake losses; the remainder was associated with liquor losses.

TABLE V-7

SUMMARY OF FILTER CAKE PROPERTIES

Parameter	Break-in Period				System Testing			
	December 1980		January 1981		February 1981		March 1981	
	Range	Average	Range	Average	Range	Average	Range	Average
Coal Sulfur Content (Wt.%)	2.8-3.6	3.3	3.0-3.9	3.3	2.6-4.2	3.5	3.3-3.9	3.6
Filter Cake Properties:[a]								
Solids Content (Wt %)	33.6-47.7	37.0	36.3-52.0	42.0	33.3-46.6	38.0	32.2-42.7	38.0
$CaSO_4/CaSO_x$(mol ratio)	0.05-0.71	0.2	0.06-0.73	0.2	0.05-0.33	0.1	0.05-0.15	0.1
Na^+/Ca (mol ratio)	0.01-0.83	0.3	0.02-1.00	0.40	0.041-0.45	0.2	0.033-0.60	0.41
Number of Displacement Wastes	---	NA[b]	---	NA[b]	---	1.5	---	0.8
Number of Spray Banks	---	2	---	2	---	2	---	2

---

[a] Values represent the average of the number of filter cake samples analyzed in the laboratory.

[b] Filter cake discharge rates not available.

Both of these problems, their causes, and impacts have already been discussed. The losses in the cake were addressed in the discussion of cake properties and liquor losses were addressed in the discussion of oxidation and sulfate precipitation.

7. Power Consumption

The power consumption by the dual alkali system is shown in Figure V-11. It ranged from 2.5% (0.53 MW) at flue gas rates equivalent to a boiler load of 21 MW, to 5.3% (0.42 MW) at an equivalent load of 8 MW.

The power consumption at Scholz was primarily related to the operation of the F.D. fan. As the pressure drop across the scrubber was typically 2 to 3 times the pressure drop across the absorber, most of the power consumed by the fan was, in turn, associated with the operation of the venturi scrubber. Thus in a typical limestone dual alkali application where flue gas is taken from an electrostatic precipitator or a baghouse and the use of a venturi is not required, one would expect the power consumption to be much lower--more in the range of 1 to 1.5% of the power generated at full boiler load.

8. Process Operability

Process operability, unlike equipment or mechanical operability, refers to the ease with which the system can be operated and controlled, to the ability of the system to adequately respond to varying conditions, and to the ability to tolerate upsets in process chemistry due to mechanical problems or operator oversight.

In general, the process operability of the limestone dual alkali system was good after the first 2 or 3 days of stable operation. It was during these initial days, following any restart of the system, that the major problems with process operability were encountered. The first 24 to 48 hours of operation with limestone feed were critical. Typically, during the first few hours of solids generation, the reactor effluent solids would settle very fast--down to 10% of the initial slurry volume in less than 5 minutes. Thereafter, as the liquor in the system was being turned over, the settling rate would steadily decline to the point where it would take 7 to 10 minutes to effect the same settling. At this point following each startup, the solids settling rate would either level off and a stable operation would be achieved or it would continue deteriorating resulting in significant solids carryover in the thickener overflow; which as discussed earlier is only conducive to the further deterioration of the quality of the solids.

10 Lines to End of Typing Area

It is possible that these restart problems were caused by the redissolution of the solids left in the thickener at the time of the outage and subsequent recrystallization into fine crystals. This appeared to be the case in the month of March when, during the eight days of downtime

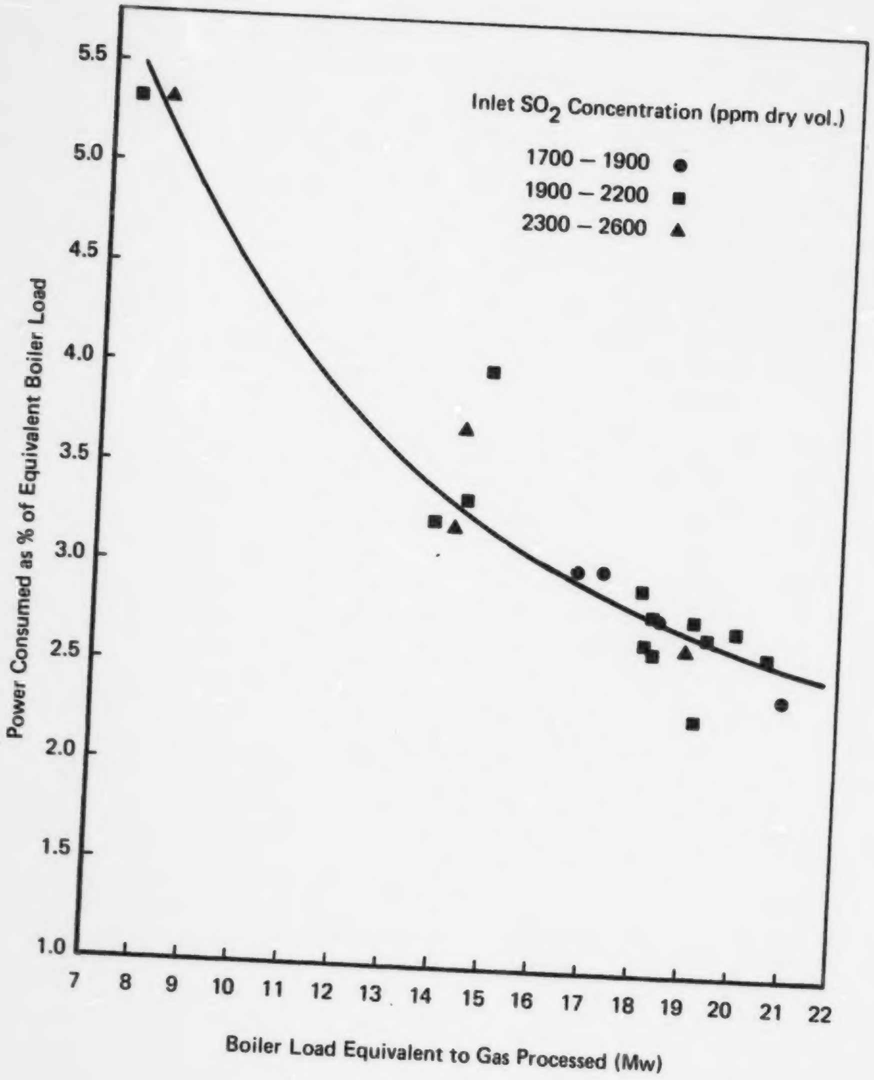


FIGURE V-11 POWER CONSUMPTION BY THE LIMESTONE DUAL ALKALI SYSTEM

for filter repairs, the solids in the thickener resuspended. This resuspension was further aggravated by the liquor circulation upon restarting the system. Thus, an increasing amount of fines were being carried over to the reactors even before generation of new solids had started.

These symptoms, however, were not necessarily present in each occurrance of the restart problem; and therefore, it appears that other causes also contributed to the problem. Further testing would be needed to identify and correct these causes.

Once the system reached stable operation, such as in the months of December 1980 or February 1981, the process operability was very good. The system was able to easily accommodate variations in inlet SO_2 concentrations of as much as 500 ppm by simply adjusting the feed forward rate of regenerated solution to the absorber/scrubber in order to maintain a constant scrubber bleed pH. Variations in boiler load and thus in the amount of gas processed by the system were also accommodated in the same fashion.

Upsets to the chemistry of the system were also handled well by the system. These upsets included the carryover of fly ash in the flue gas due to malfunction in the precipitators, the gross overfeeding of limestone due to operator oversight, and the occasional limestone and soda ash feed outages lasting from 1 to 5 hours.

Given the fact that the system was primarily controlled by the pH of the liquor, it is important to make some observations on the effect of pH variations on the process. The limestone utilization dropped considerably when the scrubber bleed pH rose above 6.0. Controlling the scrubber at a bleed pH between 5.5 and 6.0 resulted in good limestone utilization. It should be noted, however, that at scrubber bleed pH's below 5.7, the SO_2 removal efficiency drops below 92%.

The reactor train effluent pH ranged from 5.8 to 6.4. Operating the reactor train at an effluent pH below 6.0 for too long ($\sim$24 hours) causes the deterioration in the settling properties of the solids. This effect is due to two factors. First, the low pH of the regenerated liquor forces a considerable increase in the feed forward rate to the absorber/scrubber in order to maintain reasonable SO_2 removals, thus shortening the time allowed for solids to settle in the thickener. Second, the limestone utilization in the first reactor becomes very high at such low pH's. Performing such extensive regeneration in a single reactor tank favors the formation of the corresponding solids in a fine, needle-like structure, rather than in the agglomerates that exhibit good settling properties.

10 Lines
to End of
Typing Area

The highest pH observed in the reactor train effluent was 6.4, even during periods of gross limestone overfeeding. This may indicate that the process liquor becomes highly buffered at this pH. If future tests prove this to be the case, then the actual control of the regeneration step of this process by pH alone would be questionable.

Another area of concern in the SO_2 scrubbing field is the formation of scale in the system. This was not a major problem at Scholz, although some scale formation took place in the reactors and in the scrubber.

There was a slow buildup of a layer of scale on the walls of the first reactor. Insufficient agitation may have contributed to the deposition of this scale which was primarily made up of calcium sulfite/sulfate solids. The formation of this scale was also evident in the overflow pipe connecting the first and second reactors. This pipe had to be replaced in early March as the 6 inch line had been reduced to a 3-1/2 inch line due to scale buildup. This overflow line was rather long--6 feet--and had an angle of incline of only 3° because of limitations in the existing plant layout. A greater angle of incline may have reduced this scale buildup.

Another type of solids deposition in the form of round/oval beads was also observed in the first reactor. These beads--1/8 to 3/8 of an inch in size--were also made up of calcium sulfite/sulfate which built up in layers around a seed particle much like the growth of pearls. Scale buildup in the other reactors was minimal. Based on the time of operation at Scholz, it appears that a semi-annual cleaning operation as part of a regular maintenance program might be adequate to control reactor scaling.

A buildup of sodium salts took place at the top of the scrubber, near the wet/dry interface where the flue gas enters the scrubber. It is at this location that the process liquor first contacted the flue gas. A substantial evaporation of water from the scrubbing liquor took place leaving the sodium salts behind. These sodium salts are rather soluble in water and can, therefore, be easily cleaned by a periodic spray wash.

C. MECHANICAL PERFORMANCE

The mechanical performance of the equipment and its associated instrumentation represented a major source of problems at Scholz. This is evidenced by the fact that 43% of the system downtime during the startup and break-in period and 59% of the downtime during the testing period were due to mechanical problems. Not only was the availability of the process affected by these mechanical problems, but the operating conditions that could be achieved by the system were affected as well. An example of these limitations was the inability of the thickener underflow pumps to handle slurry with more than 15 wt.% solids, which required the dilution of the 20 to 25% thickener underflow slurry. Although some of the problems were caused by the limited capacity of an existing piece of equipment, most of the mechanical problems were actually due to equipment failures. The age of the equipment and its condition, even after recommissioning, were, undoubtedly, primary causes of these failures.

Some of the equipment problems have been discussed already while addressing the performance of the process. A detailed list of the equipment and instrumentation problems encountered during the program is given in Appendix E, but the most significant problems are discussed below.

1. Equipment

The most frequent or relevant equipment problems were those associated with the filter, the thickener, the raw material feeders, the pumps, and the piping.

a. Filter

The mechanical problems associated with the filter were: the low vacuum levels due to internal drum piping and automatic valve leaks; the limited filtrate handling capabilities; and the limitations in the supply of wash water. These problems have been discussed before as they affected the performance of the system in so far as waste properties were concerned.

b. Thickener

Even though the solids carryover from the thickener was primarily a process problem, it was also affected by mechanical factors. Gaps existed between the adjustable V-notched strip and the overflow weir itself. Substantial channelling of liquor took place through these gaps. Furthermore, the weir itself was not level and most of the liquor overflowed on only half of the thickener, thus not only decreasing the effective surface in the thickener but also increasing the velocity of the liquor in the areas where it did overflow. The corresponding decrease in residence time shortened the period allowed for solids settling. This was eventually corrected by using foam rubber to close the gaps and readjusting the weir to raise the low points. Another mechanical factor that affected the setttling of solids was the piping arrangement in the thickener centerwell. Both the reactor train and filtrate discharge lines were found to extend below the 3 foot deep centerwell ending only 6 to 7 feet above the thickener floor. Both discharge pipes directed the liquid straight down into the thickener. This jetting action lifted solids and otherwise disturbed the settling process. The problem went undetected until the end of January 1981 as chemical and not mechanical causes were being investigated in relation to the carryover of solids in the thickener overflow. Shortening the discharge pipes to just 2 to 3 inches below the liquid level greatly improved the flow pattern in the thickener and the clarity of the overflow.

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Typing Area

Early on in the project, another problem with the thickener was encountered during the initial startup operations when a leak developed at the bottom of the thickener. Penetrations near the support studs for the bottom plate of the thickener had occurred due to inadequate grouting. Lining separation had also taken place at various locations in the interior of the tank. This problem caused a 22 day outage.

One other minor problem with the thickener was the speed at which the rake turned since a trail of resuspended solids could be seen as the rake rotated. No action was taken to slow it down as the motor drive was a fixed speed one and a new motor drive would have been required to solve the problem.

c. Raw Material Feeders

There were problems with both the limestone and the soda ash feeders. The major problem with the soda ash feeder was its inability to feed more than 10 lbs/minute on weigh belt speed control. Normal operating conditions called for 1 to 2 lbs/minute. However, the severe sodium losses and frequent restarts imposed the need for higher feedrates. In fact, the initial startup procedure for the limestone dual alkali process called for initial feedrates of 18 to 20 lbs/minute of soda ash. This problem was not solved and operating schemes around the limitation had to be developed. Another problem was presented by lumps of soda ash getting caught in the feed gate to the weigh belt, eventually blocking the passage of soda ash altogether. Constant checking of the feeders by the operators was the only way to cope with this problem.

The limestone feed system presented an altogether different problem. Numerous pluggages took place at the base of the chute that feeds the limestone into the first reactor. Moist vapors coming up from the reactor were the cause of the problem. Although the incline in the chute (~45°) rather than a straight drop chute, also contributed to it. This problem was never corrected. A straight drop chute and the use of an air sparger to keep moist vapors from coming up the chute might alleviate the problem.

d. Pumps

The main problem associated with the operation of pumps was the need to continually replace the seals, particularly in the scrubber and thickener underflow pumps. Sometimes these pumps required new seals daily. Along with it was the need for large amounts of seal water. Two to three times as much seal water was used as was specified in the design. This was a reflection of the age and condition of the pumps. At one point in the project, a scrubber pump impeller and shaft had to be replaced due to excessive vibration of the pump shaft.

10 Lines to End of Typing Area

Another problem that has been mentioned previously was the inability of the thickener underflow pumps to pump slurries with a solids concentration greater than 15%. This limitation, due to inadequate matching of service requirements with pump capabilities during the recommissioning, required the underflow slurry be diluted with thickener underflow.

e. Piping

There were many piping problems throughout the test program. The soda ash line, in particular, was troublesome. This line delivered solution from the soda ash tank over to the thickener hold tank as well as to the thickener. The line was constructed of mostly carbon steel and could not withstand the corrosive nature of the process liquor. Near the end of the project, when higher ambient temperatures reduced the need for heat tracing, the line was replaced with a rubber hose which simply ran along the ground.

Since most of the operation took place during the winter months with unusually low temperatures, there were also many problems with frozen lines. At about 40 to 50°F, dissolved sodium salts in the liquor began to crystallize out. Thus, if the liquor in any line was allowed to drop below 40°F for any length of time, it would plug the line solid. This problem partially solved by heat tracing key lines, such as soda ash lines, thickener hold tank lines, thickener overflow lines, and thickener underflow dilution lines. Full-scale applications should provide for heat or steam tracing of all these lines.

2. Instrumentation

The most significant instrumentation problems at Scholz involved the pH units, the SO_2/O_2 monitors, the SO_2 vaporizer, and the flow controllers.

a. pH Measurement

Measuring pH is an extremely important part of controlling the process. Throughout the project, the pH probes in both the scrubber and in the last reactor were completely unreliable and were not easily accessible; and, therefore, they were difficult to clean and calibrate. As a result, pH measurements were performed manually. At least once an hour, the operators would obtain a scrubber and T-5 reactor sample and check the pH with a portable meter. This procedure was used throughout the entire test program since pH was the major process control variable. In future tests or commercial applications, it is imperative to have good, reliable pH probes that are easily accessible so they can be cleaned and calibrated. Considering that in this process pH must be controlled to within one half of a pH unit throughout the system, the accurate measurement of pH must be given careful consideration.

b. SO_2/O_2 Monitors

10 Lines to End of Typing Area

Once properly installed and calibrated, both the SO_2 and O_2 monitors did not present any significant problems, but they were not available for data collection until late in the program. Before its final installation in mid-January 1981, the SO_2 monitor required substantial field work as difficulties were encountered in calibrating the instrument, eventually necessitating the unit to be sent for service by the manufacturer. The O_2 monitor was received in January 1981 and became operational without difficulty in early February.

c. SO_2 Vaporizer

Another area of instrument problems was the control of the operation of the vaporizer. The liquid SO_2 level control would fail whenever high rates of vaporization were required. This necessitated the operators to manually fill the vaporizer every time the low level alarm came on, typically once an hour. This essentially manual operation of the vaporizer resulted in swings in the SO_2 loading to the process.

Obviously, this problem with the vaporizer control mechanism would not be a factor in a full-scale operation since no vaporizer would be used.

d. Flow Controllers

Most of the problems in controlling liquid flowrates in the system were associated with the control valves. They would stick in a given position and would not respond to the controlling signals. Again, the conditions of the valves might have been primarily responsible for these problems; nevertheless, in future systems, it might be advisable to look into the use of variable speed pumps as a means of controlling flowrates.

VI. GENERALIZED COST ESTIMATES FOR A 500 MW SYSTEM

A generalized estimate for the cost of installing and operating a limestone dual alkali system on a new 500 MW boiler is presented in this section. This estimate is based on the current knowledge and understanding of the technology. Modifications in equipment and operation may result from further testing needed prior to commercialization of this process. However, only minor changes are expected; and, therefore, this generalized estimate should be representative of the costs of commercial application of this technology.

In general, a dual alkali FGD system consists of several sub-systems for the performance of specialized functions: absorption of SO_2 from the flue gas; regeneration of the spent scrubbing solution and precipitation of waste solids; separation of the regenerated solution from the solid waste generated; storage and preparation of raw materials; disposal of the waste; and provision of utilities. Basically, the system under consideration is assumed to be designed for a 95% SO_2 removal efficiency when burning coal containing 3.5% sulfur. Additional design basis for the system is detailed in Table VI-1.

Although the dual alkali process is capable of removing particulate matter, this capability is not incorporated in the system under consideration. Thus, the absorbers are assumed to take flue gas, through booster fans, from high efficiency electrostatic precipitators. The booster fans for the dual alkali system also serve as induced draft fans for the boiler. The system is modular in nature with four absorber/two reactor trains. Each absorber is sized to handle 135 MW. Consequently, there is an 8% spare capacity at full load; and the system is capable of treating at least 80% of the design flue gas load with one absorber down. Similarly, the reactor system is capable of handling a full system load with one reactor train out of operation for short periods of time. The vacuum filters are operated and controlled independently and the number of filters in operation depends on the system load. Three filters are included--two operating and one full spare filter when operating at design conditions.

The dual alkali system is assumed to be installed in a new boiler located in the Midwest. The base year for cost estimates is mid-1980.

TABLE VI-1

DESIGN BASIS FOR THE GENERALIZED DUAL ALKALI SYSTEM

Boiler Capacity, MW	500
Boiler Location	Midwest
Boiler Status	New
Coal Heating Value, Btu/lb	12,000
Btu/kW-hr	9,000
Boiler Operating time, hrs/yr	7,000
Coal Burned, tons/yr	1,312,500
S in Coal, %	3.5
Cl in Coal, %	0.10
SO_2 in Flue Gas (94% of S in coal), lb/hr	24,675
SO_2 at Scrubber Inlet, lb/10^6 Btu	5.48
SO_2 at Stack, lb/10^6 Btu	0.274
SO_2 removal	95%
lbs/hr	23,440
tons/yr	82,044
Gas Flowrate at Scrubber Inlet, acfm	1,540,000
Temperature, °F Scrubber Inlet, acfm	310
Flue Gas Oxygen, %	6
Oxidation in Dual Alkali System, %	10
Boiler I.D. Fan and Scrubber F.D. Fan	Common
Bypass	No
Flue Gas Bypassed for Reheat, %	0
Reheat Provided by Bypass Flue Gas, °F	0
Base Year for Cost Estimation	1980
Scrubber/Reactor Trains	4/2
Spare Capacity	
Absorbers, %	8
Filters, %	50
Pumps, %	100
Instruments, Necessary for Operation, %	100
Turndown to % of Capacity	20
Sludge Disposal	Onsite
Available $CaCO_3$ in Raw Limestone, %	90%
Limestone Utilization, %	95%
Limestone Storage Pile, days	30
Soda Ash Silo Capacity, days	15

A. CAPITAL INVESTMENT

The estimated total capital investment for a generalized 500 MW limestone dual alkali system is $51.7 million, which is equivalent to $103.4/kW. This cost does not include waste cake processing or disposal. Waste processing and disposal is taken as an operating cost.

Details of this capital investment are given in Table VI-2. The process equipment and other material costs, which amount to $17.5 million, are further detailed in Table VI-3. These material costs are based on the actual costs of a 300 MW lime dual alkali system installed in Unit No. 6 of Louisville Gas and Electric's Cane Run Station[1], and on estimated costs by Tennessee Valley Authority (TVA)[2]. The difference in number of modules due to size (500 MW vs. 300 MW) and the difference in reactor configuration and in raw material feed units between lime- and limestone-based dual alkali systems are taken into consideration.

Common fans are used for the dual alkali system and the boiler. One fan is provided for each of the four absorbers. The fan cost for the dual alkali system is the incremental cost over that for each induced draft fan required for the operation of the boilers. The incremental cost is equal to the difference in cost between a fan used for a pressure drop of 27 inches WG and one used for 15 inches WG. The absorbers are equipped with spray nozzles for presaturation of the gas and trays for promoting gas/liquid contact.

Each of the two reactor trains has four agitated reactor tanks. Each reactor overflows sequentially into the next reactor. The last reactor, which overflows to the thickener, is also equipped with transfer pumps to drain the reactor system.

The soda ash silo has a storage capacity of 15 days for operation at design conditions. It is equipped with vibrating bottoms, weighing and feeding mechanisms. The limestone is first conveyed to a storage pile and from the pile to a day bin. The limestone in the bin is first crushed and then processed by a wet ball mill to reduce its size, prior to being fed to the system.

The cost of erecting the system is estimated at 50% of the process equipment and additional material costs. The service, utilities, and miscellaneous costs, taken as 6% of all other direct costs, cover items

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[1] Van Ness, R.P., et al., "Full-Scale Dual-Alkali Demonstration at Louisville Gas and Electric Co.: Final Design and System Cost," EPA-600/7-79-221b, September 1979.

[2] Tomlinson, S.V., et al., "Definitive SO_x Control Process Evaluations: Limestone Double Alkali and Citrate FGD Processes," EPA-600/7-79-177, August 1979.

TABLE VI-2

ESTIMATED CAPITAL INVESTMENT FOR A 500 MW LIMESTONE DUAL ALKALI SYSTEM

(1980 Dollars)

	(In Thousands)
• Direct Costs	
Process Equipment and Material Costs	$17,510
Erection Costs	8,760
Services, Utilities, and Miscellaneous	1,580
Total Direct	$27,850
• Indirect Costs	
Engineering Design (9% of D.C.)	2,510
Construction Field Expense (11% of D.C.)	3,060
Contractor's Fee (5% of D.C.)	1,390
Contingency (20% of D.C.)	5,570
Total Indirect	$12,530
• Allowance for Startup and Modification	4,040
• Interest During Construction	4,850
Total Depreciable	$49,270
• Allowance for Spare Parts	350
• Land	810
• Working Capital	1,240
Total Investment	$51,670

TABLE VI-3

ESTIMATED COST OF PROCESS EQUIPMENT AND MATERIALS FOR A 500 MW LIMESTONE DUAL ALKALI SYSTEM

(1980 Dollars)

	(In Thousands)
Major Pieces of Equipment	
Absorbers	$ 2,020
Agitators	270
Ball Mill and Crusher	470
Fans and Fluid Drives	960
Miscellaneous Handling and Dust Collecting	340
Pumps and Motors	1,410
Reheaters	300
Tanks and Silos	1,750
Thickener Shell and Mechanism	540
Vacuum Filters	1,600
Weigh Feeders and Conveyors	250
	$ 9,910
Additional Material Costs	
Buildings	250
Ducting	1,710
Electrical and Instrumentation	1,860
Foundations and Structural Steel	1,320
Land Improvements and Miscellaneous	250
Lining	710
Piping and Insulation	1,500
	$ 7,600
	$17,510

such as maintenance shops, analytical laboratories, communication, security, and office-related costs; walkways; landscaping, etc. The indirect costs are estimated in direct proportion to the total direct costs: engineering at 9%; construction field expense at 11%; contractor's fee at 5%; and contingency at 20%.

Based on TVA's estimates, the allowance for startup and modification is taken at 10% of total direct and indirect costs; and the interest during construction is taken at 12% of total direct and indirect costs.

The allowance for spare parts is estimated at 2% of the process equipment and additional materials. The cost of land given in Table VI-2 includes the land needed for the FGD system and for waste disposal, which is estimated at 210 acres. It is assumed that the expansion of the disposal site will be made progressively with the operation of the plant so the cost of land development is treated as an operating cost. The working capital requirement is estimated based on TVA's guidelines: three weeks of raw materials, seven weeks of other direct costs, and seven weeks of overhead.

B. OPERATING COSTS

The estimated annual operating costs for a 500 MW system are given in Table VI-4. The average labor is estimated at three operators per shift, three shifts per day. The cost of one supervisor and one analyst are included in the labor costs. The plant overhead is estimated at 50% of the operating and maintenance costs. Administrative overhead is estimated at 10% of operating and maintenance labor. Maintenance materials are estimated at 5% of direct capital investment. Capital charges, including interest, depreciation, return on investment, taxes, and insurances are taken at 14.6% of the total fixed plant investment.

The total annual operating costs (including capital charges) amount to \$17.7 million, equivalent to 5.1 mills/kWh or \$216/ton SO_2 removed.

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TABLE VI-4

ESTIMATED ANNUAL OPERATING COSTS FOR A 500 MW LIMESTONE DUAL ALKALI SYSTEM

95% SO_2 Removal
(1980 Dollars)

Direct Costs	Annual Quantity	Unit Cost $	Annual Cost $
Raw Materials			
Limestone	143,990 tons	12/ton	$ 1,727,900
Soda Ash	5,440 tons	110/ton	598,400
			$ 2,326,300
Utilities			
Electricity	41.1×10^6 kw-hr	0.035/kw-hr	1,438,500
Water	324×10^6 gals	0.20/1000 gal	64,800
Steam	400×10^6 lbs	2.0/1000 lbs	800,000
			$ 2,303,300
Labor and Maintenance			
Operating Labor	3 shifts	120,000/shift	360,000
Supervision	2,080 hours	20/hour	41,600
Analysis	2,080 hours	15/hour	31,200
Maintenance	5% of direct investment		1,392,500
			1,825,300
Sludge Disposal	3.15×10^5 wet ton	10/wet ton	3,150,000
TOTAL DIRECT COSTS			$ 9,604,900
Indirect Costs			
Capital Charges	14.6% of Depreciable Investment		7,193,400
Overhead			
Plant	50% of Labor and Maintenance		912,700
Administrative	10% of Operating Labor		36,000
TOTAL INDIRECT COST			$ 8,142,100
TOTAL ANNUAL OPERATING COST			$17,747,000
Mills/kWh			5.1
$/ton SO_2 Removed			216

Basis: Operating time 7,000 hr/yr
3.5% S in coal
Reheat - 35°F
Average SO_2 Removal - 391 lbs/min
Onsite Solids Disposal by Trucking

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VII. CONCLUSIONS

A final and complete evaluation of the limestone dual alkali technology is not possible at this time because of the limitations encountered in the performance of this test program. Nevertheless, a greal deal was learned about the process and some conclusions can be drawn from this experience.

The conclusions and recommendations given herein are based on the following facts:

- The testing of the system was significantly reduced because of budgetary considerations;
- The SO_2 removal efficiency of the limestone dual alkali system at Scholz proved to be excellent;
- Limestone utilization by the system was also excellent;
- There were periods when solids with excellent settling characteristics were generated; there were periods when the solids exhibited very poor settling characteristics;
- The waste filter cake was too moist and sodium losses in the cake were too high;
- Sodium losses in the cake were reduced to better than design specifications whenever the proper amount of wash water was used;
- The overall soda ash consumption was excessive; and
- The mechanical performance of the equipment and instrumentation was poor and contributed to some of the above problems.

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These facts lead to the following conclusions:

- The limestone dual alkali process appears to be technically feasible; and
- Further testing is required to reinforce the above conclusion and to develop sufficient process information needed for full-scale commercialization purposes.

It is also recommended that laboratory or small pilot plant tests, to better understand the generation of solids with good settling characteristics, be performed prior to any further prototype scale testing.

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APPENDIX A

PROCESS AND BOILER OPERATING SUMMARIES

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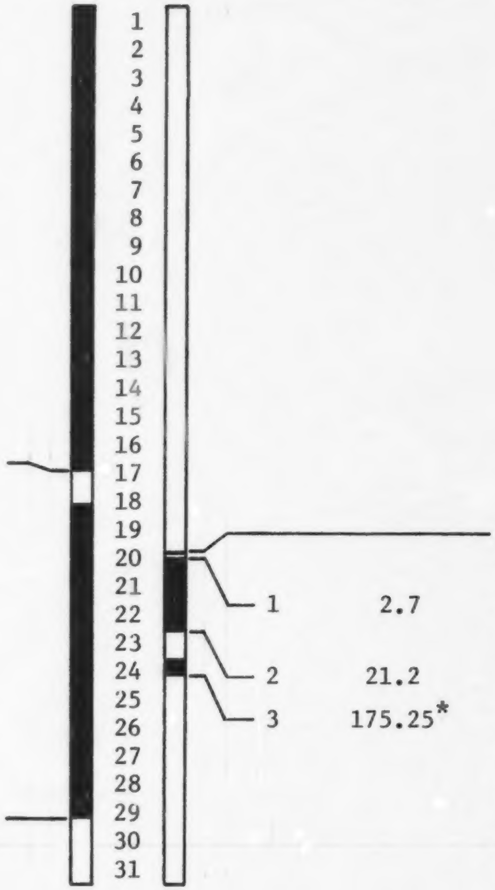
LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - AUGUST 1980

Boiler - Unit No. 1 - Outage

Cause	Duration (hours)	No.
Economizer Tube Leak	33.1	1
Load Orders	52.5*	2



FGD System Outage

No.	Duration* (hours)	Cause
		First admission of flue gas to the system
1	2.7	Removing debris from the gas duct that was accumulating in the scrubber
2	21.2	Rupture in the scrubber recycle line
3	175.25*	Bottom of the thickener tank leaking. Repairs required repairing the damaged steel, relining the bottom of the tank and affected thickener wall areas, and grouting underneath the tank.

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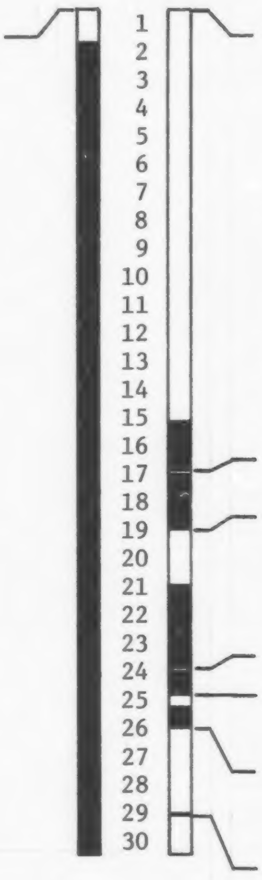
*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - SEPTEMBER 1980

Boiler - Unit No. 1 - Outage

Cause	Duration (hours)	No.
See outage 2 - August	29.4*	2



FGD System Outage

No.	Duration (hours)	Cause
3	354.2*	See outage 3 - August
4	1.1	Inspect hole in scrubber wall
5	46.8	Increase active sodium concentration
6	0.1	Gulf Power checking their SO_2 analyzer
7	10.7	Fix level control valves in absorber and scrubber
8	74.4	Poor settling solids. Increasing active sodium and fixing automatic valve in filter
9	31.6	Excessive fan vibration

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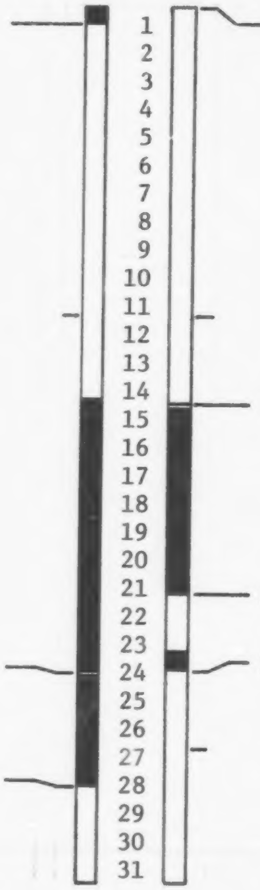
*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - OCTOBER 1980



Boiler - Unit No. 1 - Outage				FGD System Outage		
Cause	Duration (hours)	No.	Day	No.	Duration (hours)	Cause
Load Orders	321.3	3	1	9	335.6*	See outage 9 - September. Balancing fan; fixing hole in scrubber; resurfacing bottom of reactor T-1, repair thickener rake lifting mechanism
Boiler water preheater problems			12			Gulf Power Unit 1 down
			15	10	2.4	Removing duct debris from scrubber
			21	11	47.75	Poor settling solids
Leak in heater relief valve	1.2	4	24	12	181.5*	Repairing broken end of filter drum. Need to remove poor settling solids
Load orders	82.6	5	28			Legal issues associated with CEA's bankruptcy declaration being resolved

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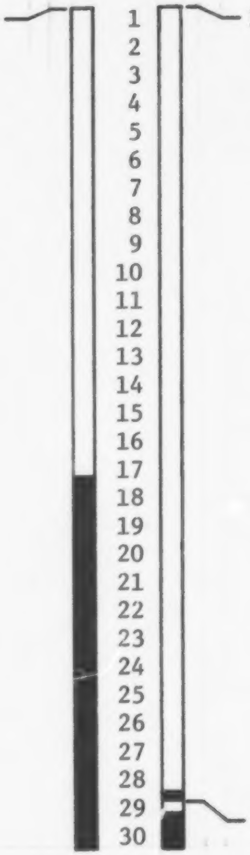
*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - NOVEMBER 1980



Boiler - Unit No. 1 - Outage			FGD System Outage		
Cause	Duration (hours)	No.	No.	Duration (hours)	Cause
See outage 5 - October	403.1*	5	12	671.6*	Legal issues associated with CEA's bankruptcy declaration being resolved.
					Fixing leaky valves, flanges, and pumps; repairing soda ash feeder; installing SO_2 injection system; repairing broken end of filter drum and fixing filter automatic valve
			13	13.6	Leaky control valve in reactor T-5. Plugged lime from reactor to the thickener

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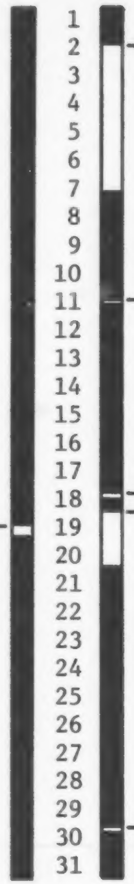
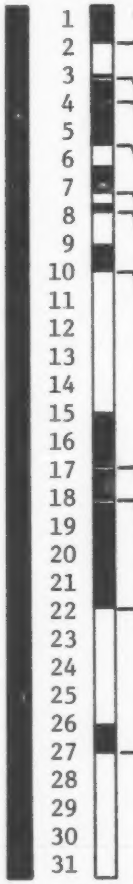
*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - DECEMBER 1980



Boiler - Unit No. 1 - Outage				FGD System Outage		
Cause	Duration (hours)	No.	Day	No.	Duration (hours)	Cause
			1			
			2	14	126.3	Out of soda ash
			3			
			4			
			5			
			6			
			7			
			8			
			9			
			10			
			11	15	1.4	Leak in feed forward flow control valve
			12			
			13			
			14			
			15			
			16			
			17			
			18	17	2.9	Rupture in scrubber bleed line
Air heater plugged	10.1	6	19	18	46.75	Out of soda ash
			20			
			21			
			22			
			23			
			24			
			25			
			26			
			27			
			28			
			29			
			30	19	2.0	Suction port to reactor T-5 pump plugged
			31			

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - JANUARY 1981



Boiler - Unit No. 1 - Outage			FGD System Outage		
Cause	Duration (hours)	No.	No.	Duration (hours)	Cause
			20	33.5	Leak in thickener hold tank pump. Back-up pump plugged. Fix leaks in piping, pumps, and valves
			21	1.5	Cloudy thickener
			22	0.2	Welding scrubber pH slipstream line
			23	19.5	Solids in thickener overflow
			24	6.5	Solids in thickener overflow
			25	30.0	Solids in thickener overflow
			26	120.2	Solids in thickener overflow. Allow settling of solids and filter them out. Extensive freeze prevention, heat tracing, instrument protection "Thawing" the plant, thawing frozen valves and pipes, reinstalling flow meters, etc.
			27	0.4	Scrubber pump failure
			28	0.9	Limestone chute plugged
			29	102.25	Excessive amount of solids in thickener overflow. Allow settling of solids and filter them out.
			30	105.8*	Excessive amount of solids in thickener overflow. Correcting return lines to centerwell They were found to extend as far down as the centerwell itself. Leveling and sealing gaps in thickener overflow weir.

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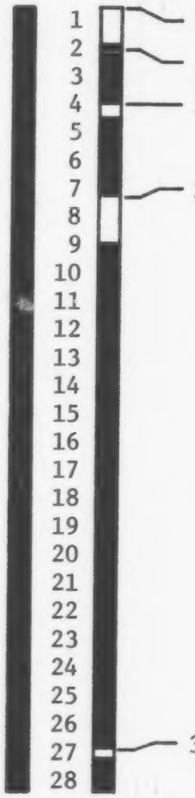
*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - FEBRUARY 1981



Boiler - Unit No. 1 - Outage				FGD System Outage		
Cause	Duration (hours)	No.	Day	No.	Duration (hours)	Cause
			1	30	36.9*	See outage 30 - January
			2	31	0.2	Fan maintenance
			4	32	11.25	Cloudy thickener
			7	33	38.75	Roads to waste disposal area washed out by heavy rains. Unable to remove any more solids from system.
			27	34	6.0	Both scrubber pumps out of service

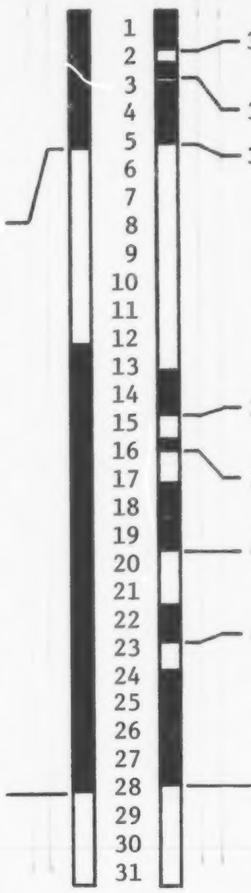
---

*This month only

LIMESTONE DUAL ALKALI SYSTEM
SCHOLZ STEAM PLANT
OPERATING SUMMARY - MARCH 1981

Boiler - Unit No. 1 - Outage

Cause	Duration (hours)	No.
Load Orders	168.25	7
Load Orders		



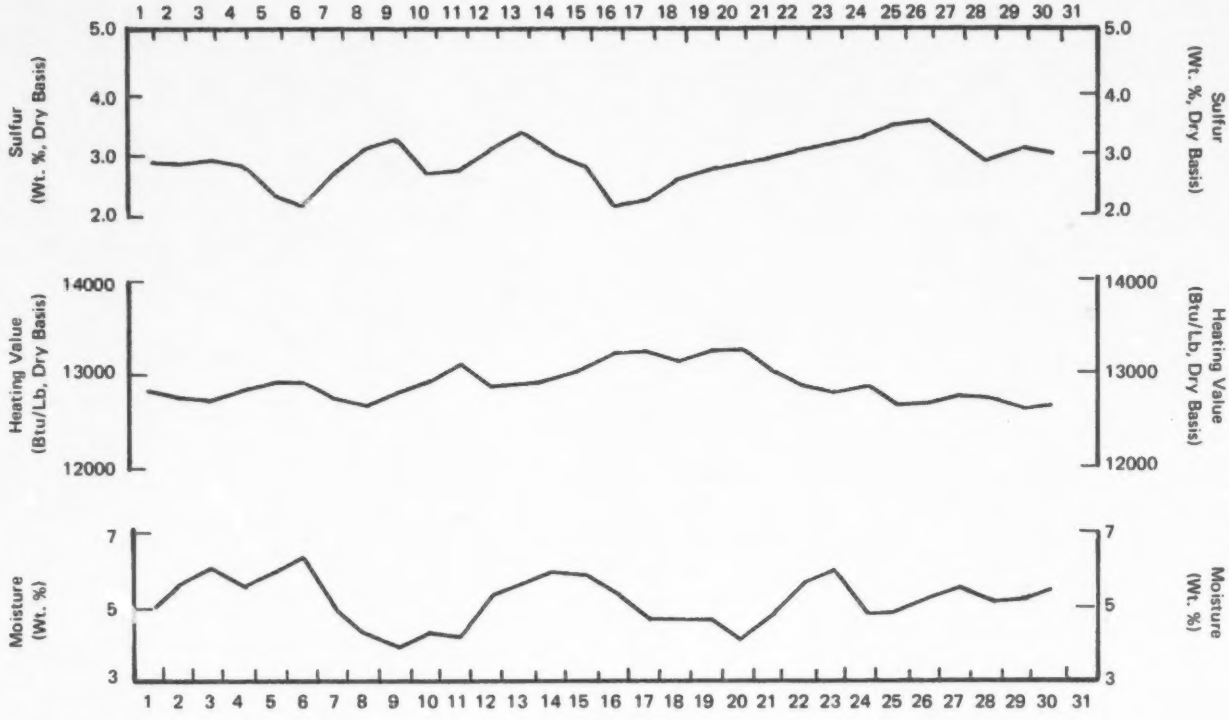
FGD System Outage

No.	Duration (hours)	Cause
35	6.5	Replace impeller in scrubber pump Motor in back-up pump inoperative
36	1.25	Leaky packing in scrubber pump
37	190.3	Internal piping in the drum of the vacuum filter being rebuilt. Leaks in the piping had rendered half of the filter drum surface inoperative
38	19.5	Resuspension of solids in the thickener
39	28.3	Resuspension of solids in the thickener
40	45.75	Poor settling solids being generated in reactors
41	22.7	Rake locked in place
		Conclusion of system testing

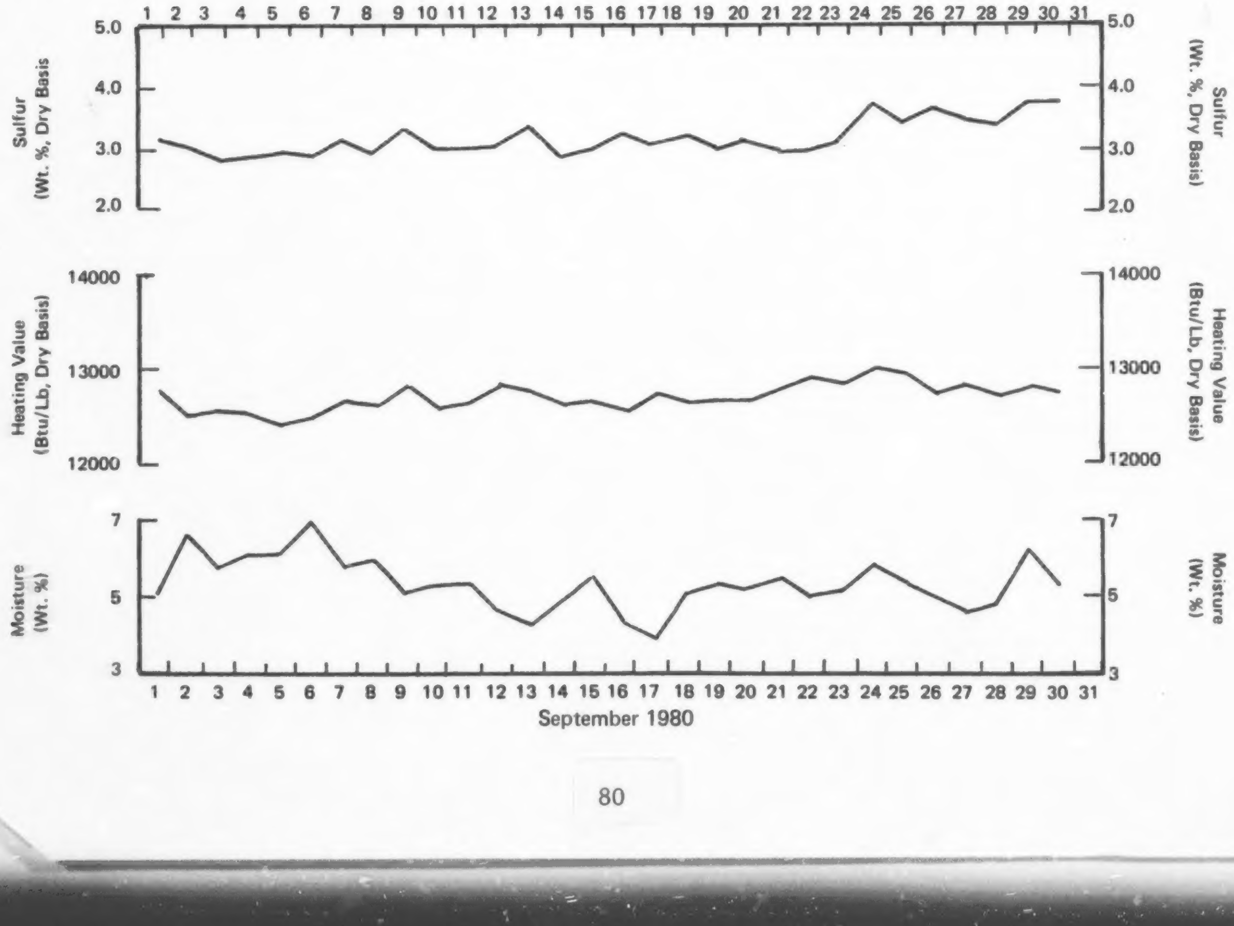
APPENDIX B

DAILY COAL ANALYSIS

DAILY COMPOSITE OF COAL FIRED IN UNIT 1

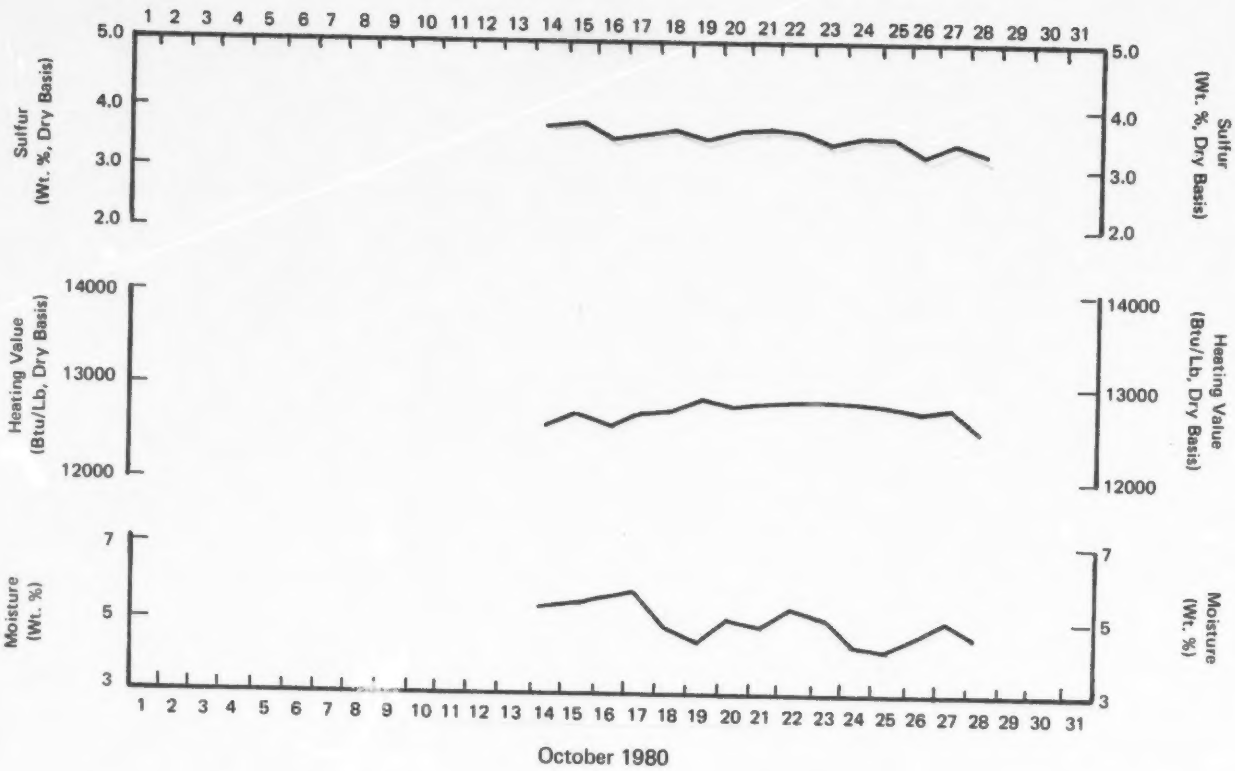


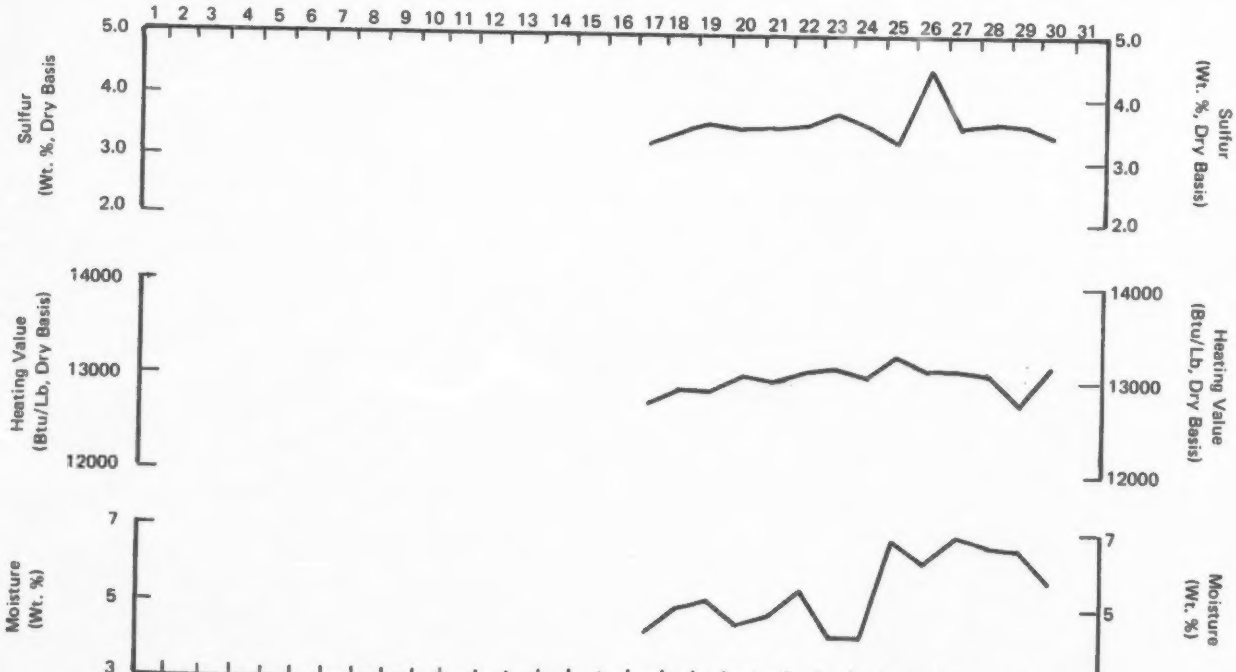
August 1980



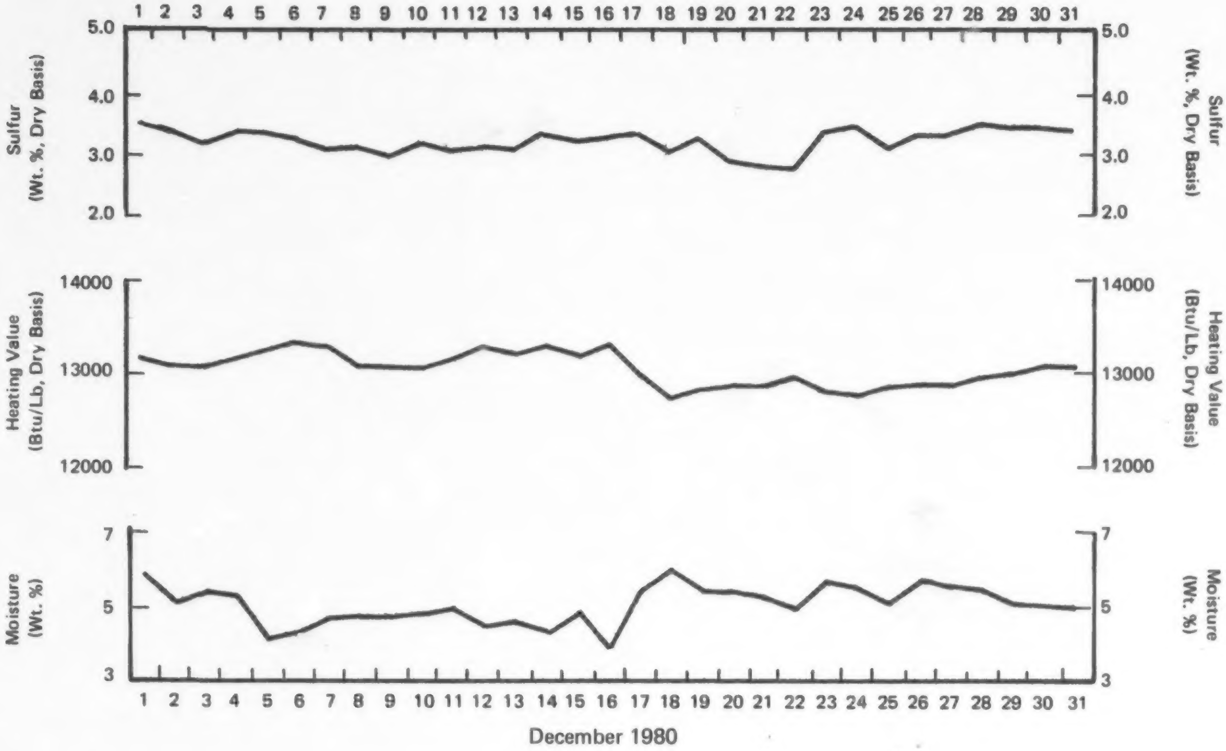
September 1980

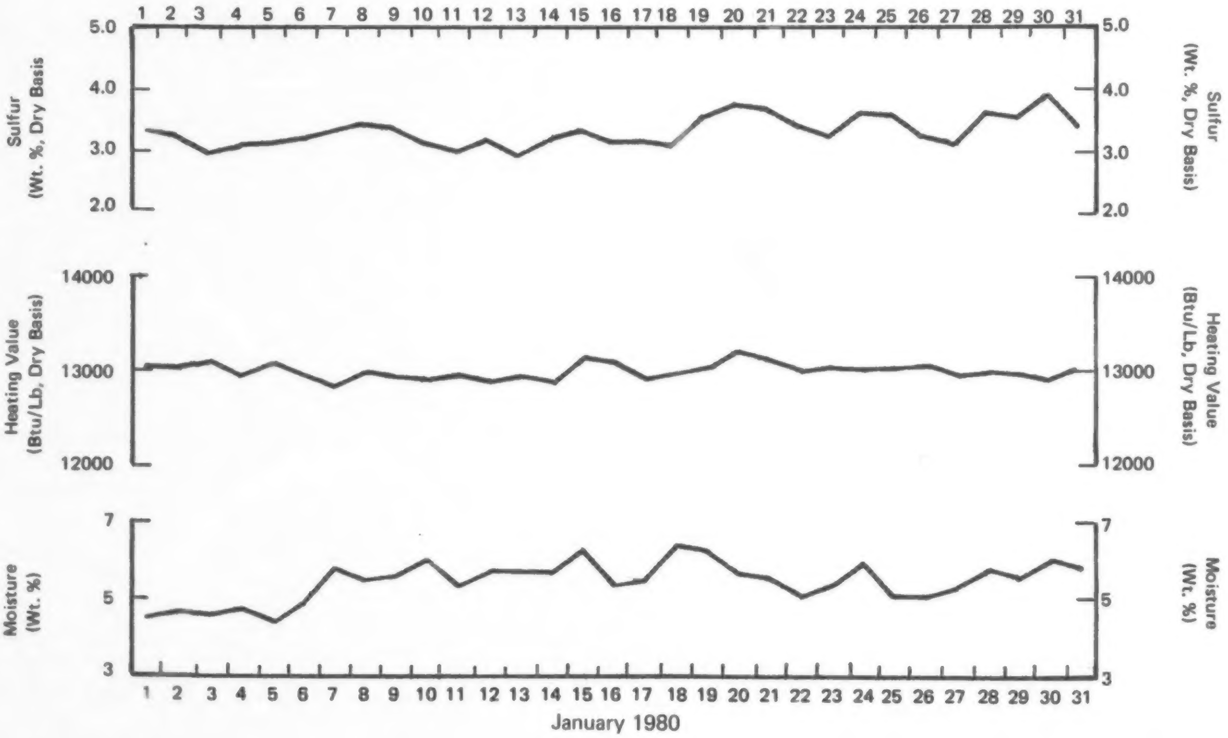
DAILY COMPOSITE OF COAL FIRED IN UNIT 1



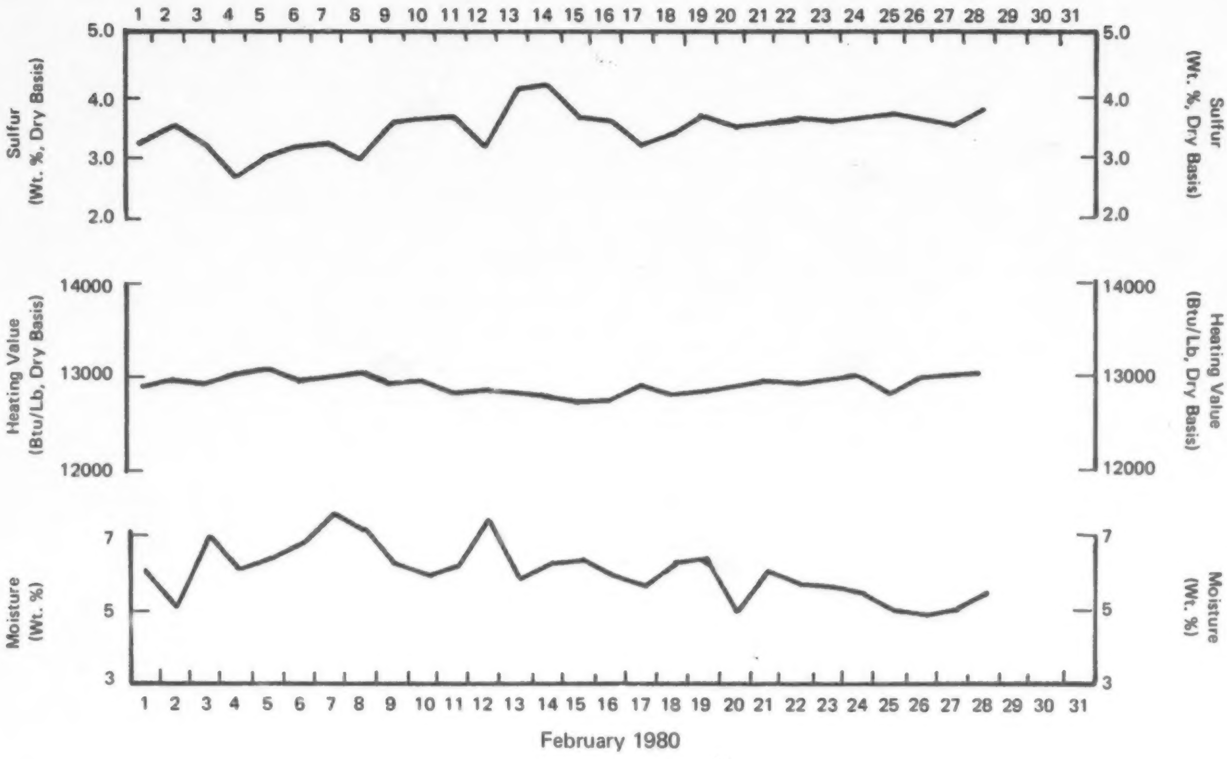


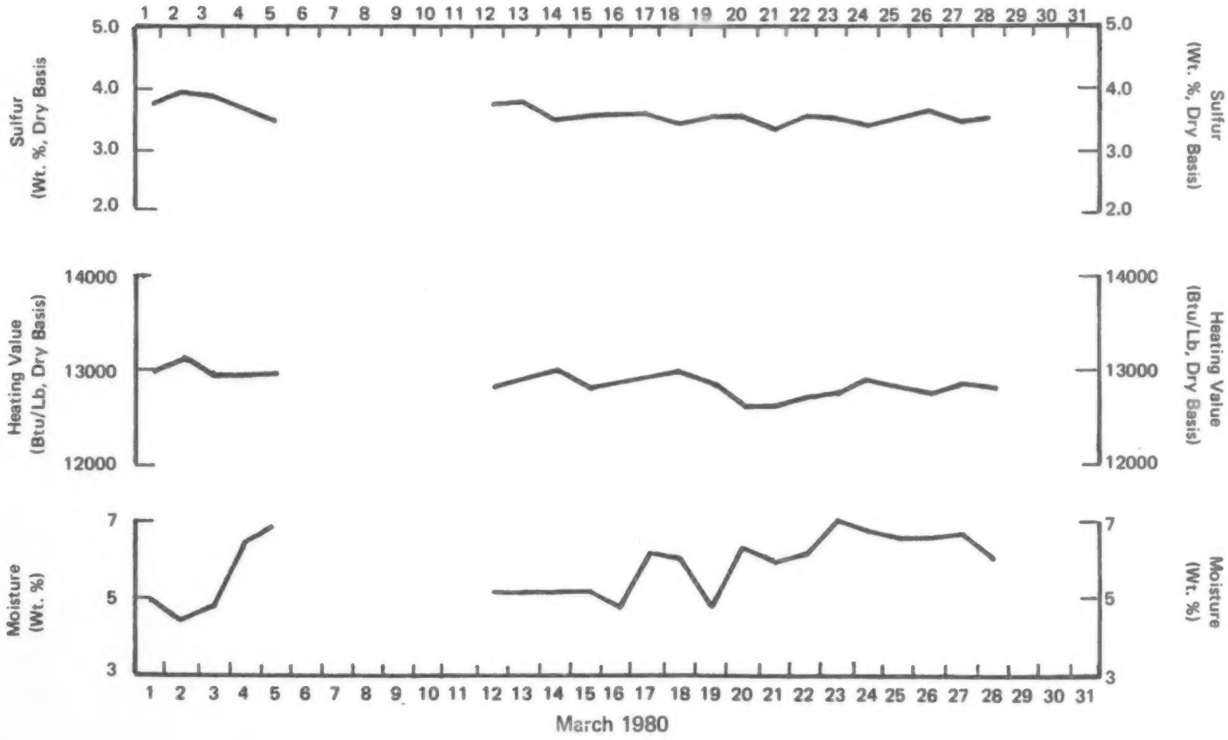
DAILY COMPOSITE OF COAL FIRED IN UNIT 1





DAILY COMPOSITE OF COAL FIRED IN UNIT 1



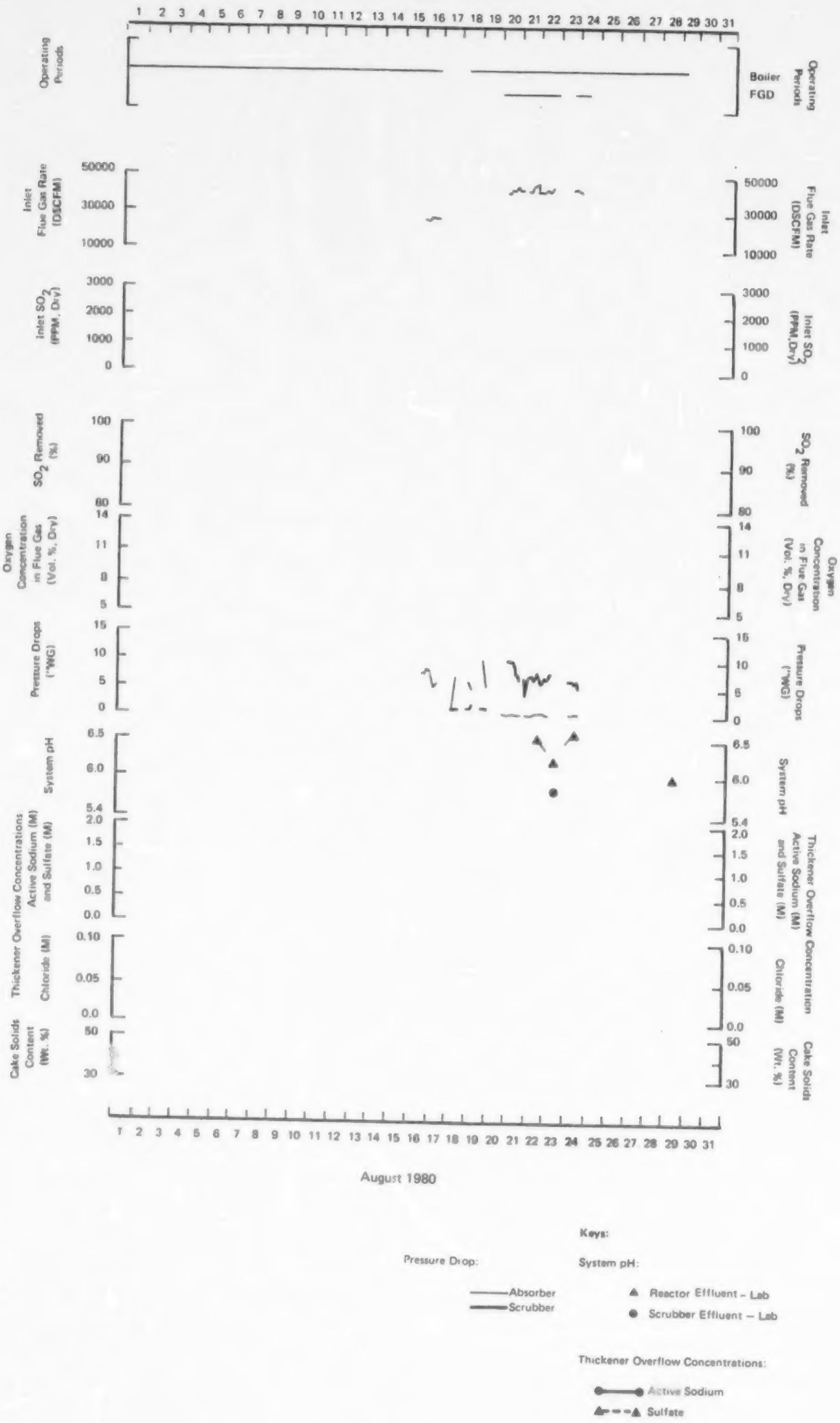


APPENDIX C

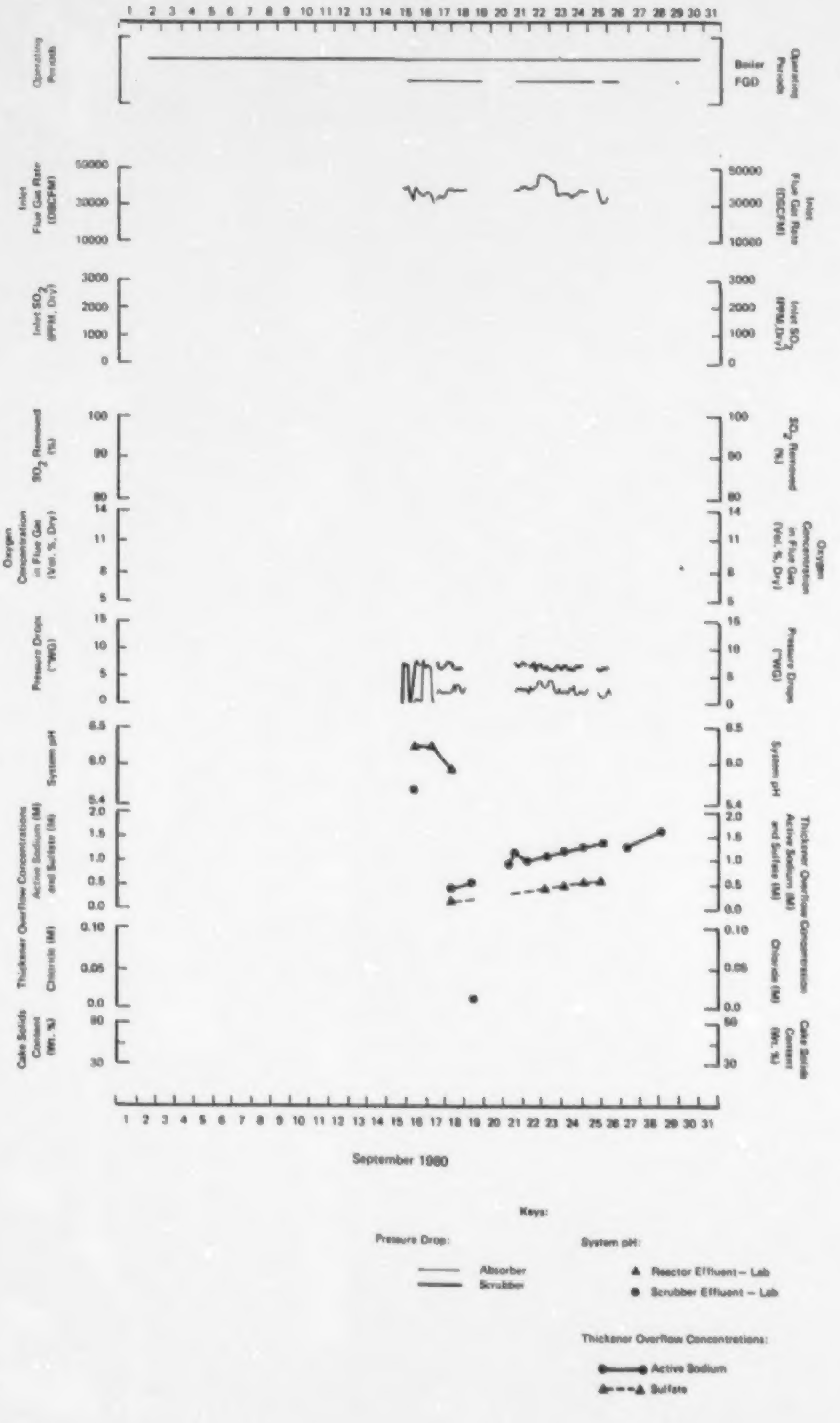
OPERATING CONDITIONS - TIME CHARTS

10 Lines
to End of
Typing Area

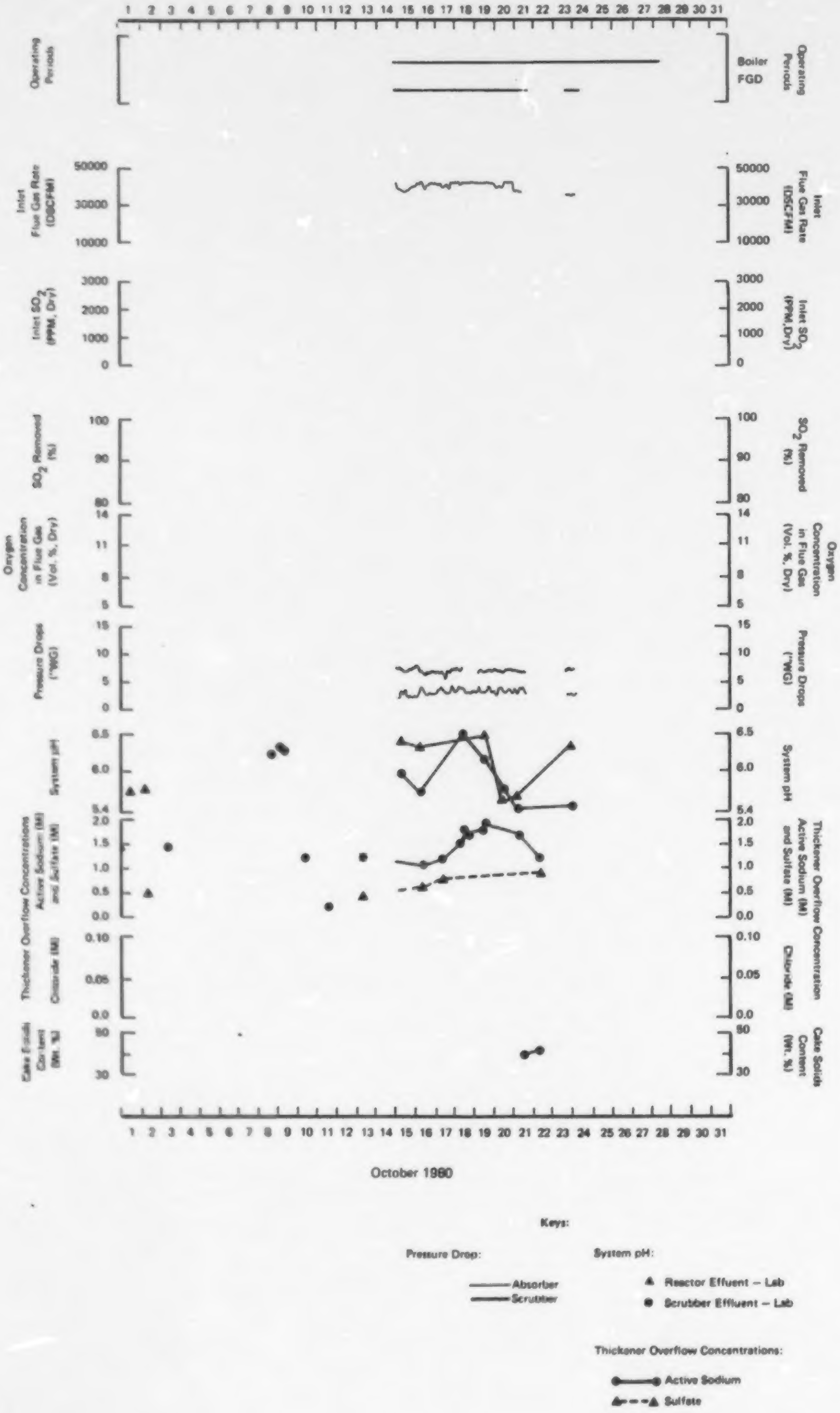
OPERATING DATA



OPERATING DATA



OPERATING DATA



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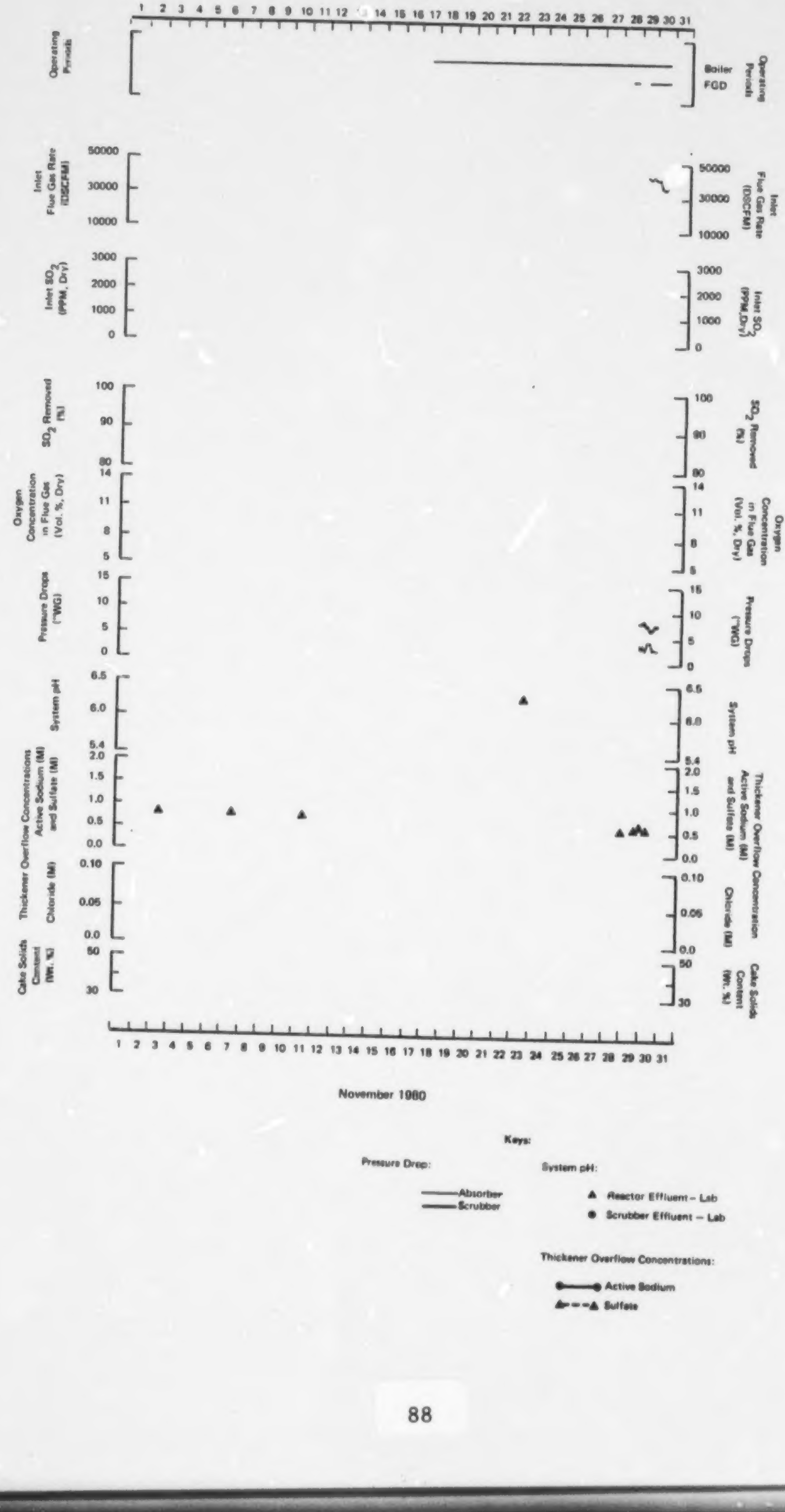

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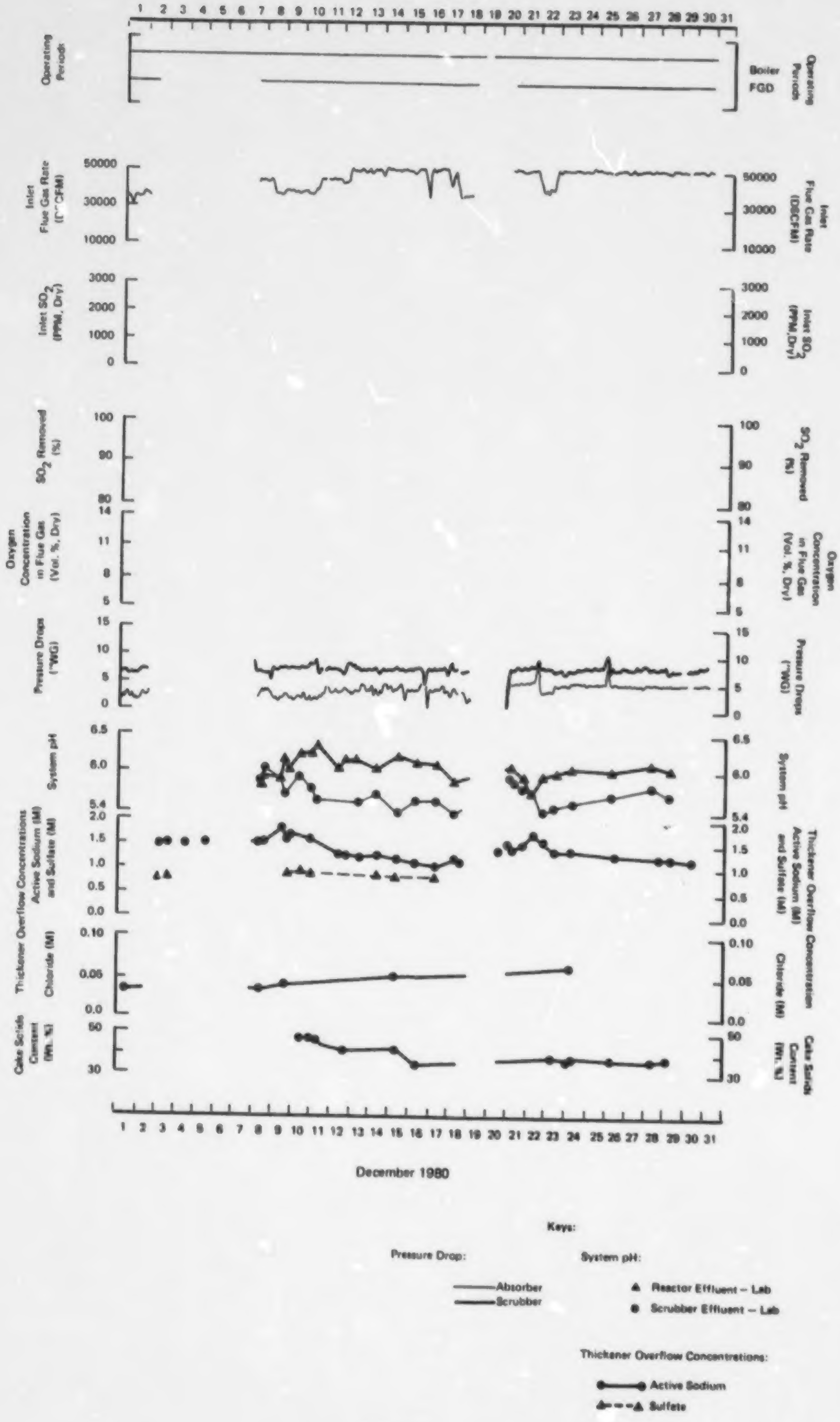

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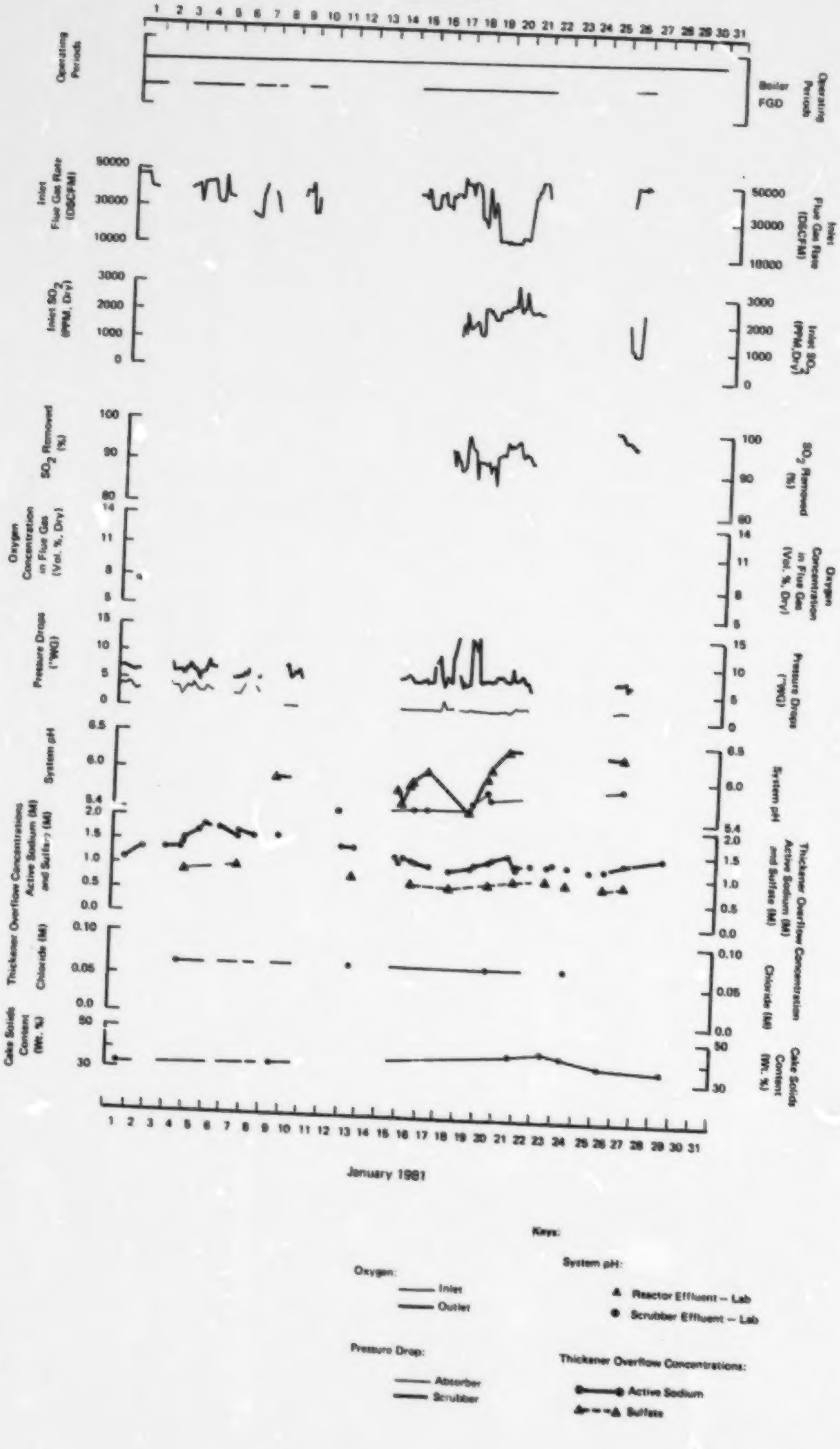

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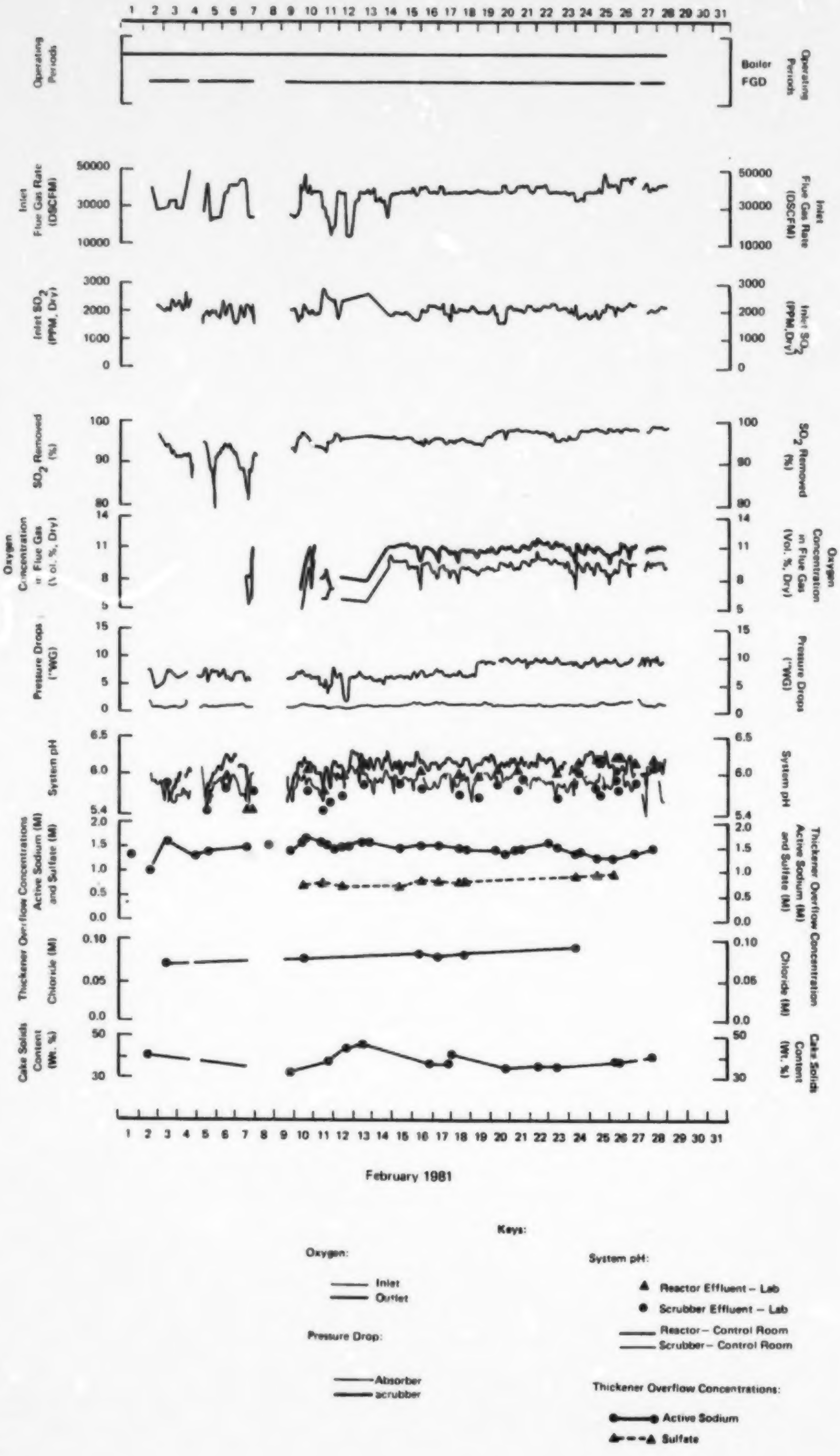

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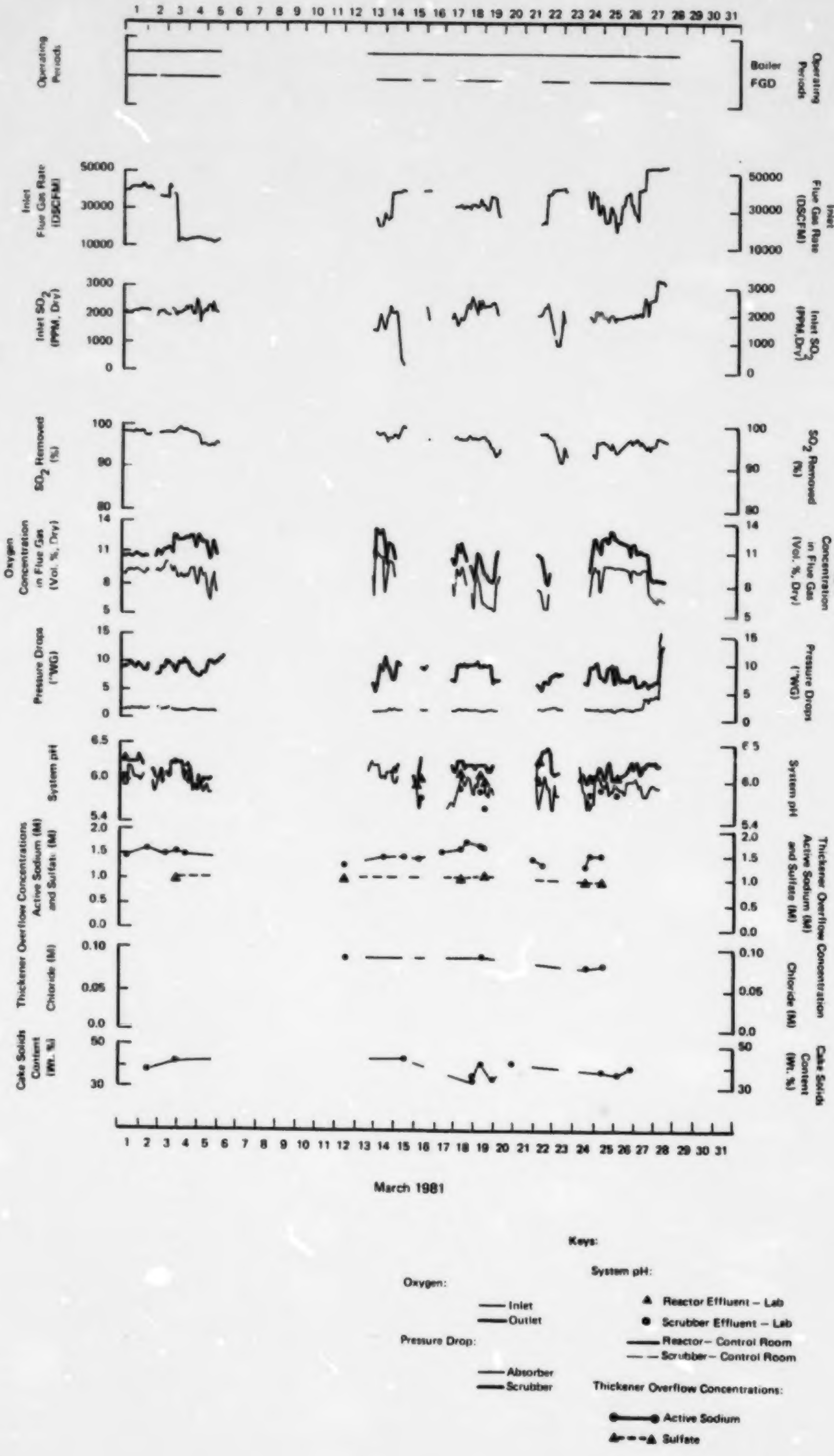


APPENDIX D

TYPICAL PROCESS FLOWS AND STREAM COMPOSITIONS

10 Lines
to Foot of
Typing Area

APPENDIX D

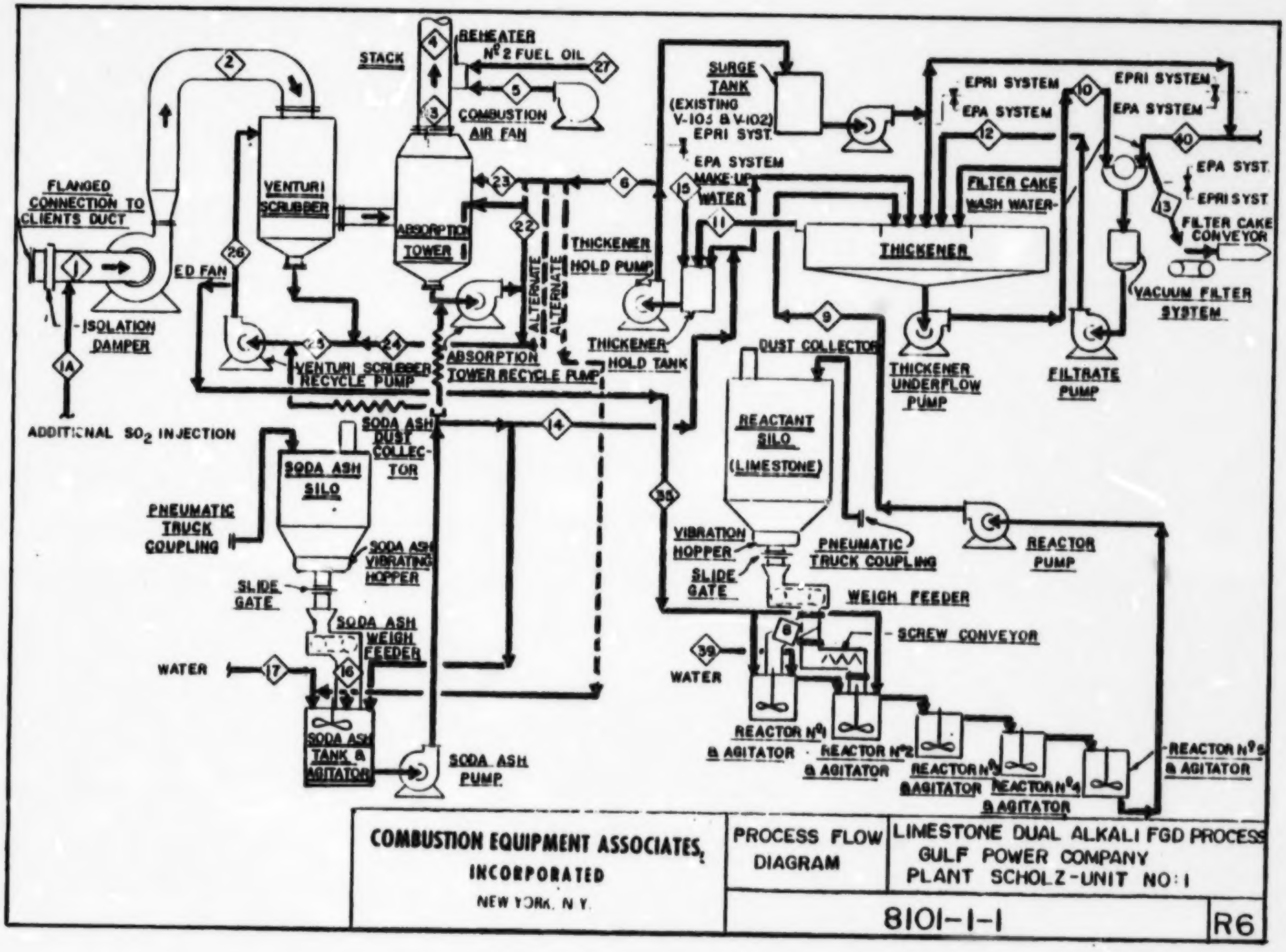
TYPICAL PROCESS FLOWS AND STREAM COMPOSITIONS

The three system compositions shown in this appendix were selected because of the extremely different process conditions they represent.

- The stream compositions on February 18, 1981 were obtained when the system was in an excellent operating mode. The system had been running in a stable mode for nine days without interruptions. The system compositions and performance on this day were typical of that for the month of February.
- On March 26, 1981, the process was in a fair operating mode. Process data and liquor and solids samples were collected just as the solids settling characteristics were beginning to deteriorate.
- The characterization of the system on March 27, 1981 was made at a time when the settling characteristics of the solids had completely deteriorated.

The stream numbers correspond to the enclosed process flow diagram.

10 Lines
to End of
Typing Area


STACK
REHEATER
Nº2 FUEL OIL
COMBUSTION AIR FAN
FLANGED CONNECTION TO CLIENTS DUCT
ED FAN
ISOLATION DAMPER
ADDITIONAL SO2 INJECTION
VENTURI SCRUBBER
VENTURI SCRUBBER RECYCLE PUMP
ABSORPTION TOWER
ABSORPTION TOWER RECYCLE PUMP
ALTERNATE
ALTERNATE
THICKENER HOLD PUMP
THICKENER HOLD TANK
SURGE TANK
(EXISTING V-105 & V-102)
EPRI SYST.
EPA SYSTEM
MAKE-UP WATER
EPRI SYSTEM
EPA SYSTEM
EPRI SYSTEM
EPA SYSTEM
EPA SYST.
EPRI SYST.
FILTER CAKE WASH WATER
THICKENER
FILTER CAKE CONVEYOR
VACUUM FILTER SYSTEM
THICKENER UNDERFLOW PUMP
FILTRATE PUMP
DUST COLLECTOR
REACTANT SILO (LIMESTONE)
VIBRATION HOPPER
SLIDE GATE
PNEUMATIC TRUCK COUPLING
WEIGH FEEDER
SCREW CONVEYOR
REACTOR PUMP
WATER
REACTOR Nº1 & AGITATOR
REACTOR Nº2 & AGITATOR
REACTOR Nº3 & AGITATOR
REACTOR Nº4 & AGITATOR
REACTOR Nº5 & AGITATOR
SODA ASH DUST COLLECTOR
SODA ASH SILO
PNEUMATIC TRUCK COUPLING
SODA ASH VIBRATING HOPPER
SLIDE GATE
SODA ASH WEIGH FEEDER
WATER
SODA ASH TANK & AGITATOR
SODA ASH PUMP
1A
1
2
3
4
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40

COMBUSTION EQUIPMENT ASSOCIATES,
INCORPORATED
NEW YORK, N.Y.

PROCESS FLOW DIAGRAM

LIMESTONE DUAL ALKALI FGD PROCESS
GULF POWER COMPANY
PLANT SCHOLZ-UNIT NO:1

8101-1-1

R6

TABLE D-1

STREAM COMPOSITION - FEBRUARY 18, 1981

	Gas Stream					Feed Stream			
Stream Number	2	--	3	5	4	6	16	40	15
Location	Inlet Flue Gas	Scrubber Exit	Absorber Exit	Reheater Air	Stack Gas	Limestone Feed	Soda Ash Feed	Wash Water	Make-up H_2O
Temperature (°F)	286	131	118	--	113	--	--	--	--
Flow Rate (dscfm or gpm* or lbs/min**)	38,480	--	--	6,410	44,890	15.0 **	3.0 **	10.0 *	0.0 *
SO_2 (ppm dry)	2,050	--	99.2	--	85.0	--	--	--	--
O_2 (vol.% dry)	9.0	--	--	21.0 ‡	10.7	--	--	--	--
Pressure Drop ("WG)	-0.2	6.5	1.7	--	--	--	--	--	--
Composition									
Percent Suspended Solids	--	--	--	--	--	99.95	99.92	--	--
Solids Analysis (mmoles/gm)									
Ca	--	--	--	--	--	9.63	--	--	--
Mg	--	--	--	--	--	0.123	--	--	--
Na	--	--	--	--	--	--	18.82	--	--
Fe	--	--	--	--	--	0.004	--	--	--
CO_3	--	--	--	--	--	9.75	9.41		

	Process Liquor and Slurry Streams													
Stream Number	6	23	22	24	26	35	--	--	--	--	9	10	13	11
Location	Absorber Feed	Tray Feed	Absorber Recycle	Absorber Bleed	Scrubber Recycle	Scrubber Bleed	T-1 Bleed	T-2 Bleed	T-3 Bleed	T-4 Bleed	T-5 Bleed	Thickener Underflow	Filter Cake	Thickener Overflow
Temperature (°F)	NA[a]	NA[a]	120	120	124	124	124	123.5	123	123	123	--	--	NA[a]
Flow Rate (gpm)	150	150	2,570	LC[b]	1,190	138	GF[c]	GF[c]	GF[c]	GF[c]	LC[b]	--	--	LC[b]
Composition														
Percent Suspended Solids	--	--	--	--	--	--	1.17	1.36	1.39	1.42	1.51	16.4	34.9	0.025
Liquid Analysis (M)														
pH	--	--	--	--	--	--	--	--	--	--	--	--	--	--
HSO_3^-	0.568	--	--	0.673	--	0.924	0.867	0.789	0.763	0.727	0.720	--	--	0.629
$SO_3^=$	0.433	--	--	0.362	--	0.306	0.322	0.371	0.391	0.412	0.416	--	--	0.422
$SO_4^=$	0.837	--	--	0.843	--	0.923	0.913	0.912	0.912	0.909	0.909	--	--	0.836
HCO_3^-	0.0.2	--	--	0.011	--	0.010	0.010	0.010	0.010	0.010	0.010	--	--	0.009
H_2CO_3	0.024	--	--	0.023	--	0.021	0.021	0.021	0.021	0.021	0.021	--	--	0.018
Cl^-	--	--	--	--	--	0.090	--	--	--	--	--	--	--	0.079
Mg^{++} (ppm)	645.0	--	--	653.0	--	093.0	751.0	685.0	700.0	693.0	713.0	--	--	678.0
Ca^{++} (ppm)	15.1	--	--	14.0	--	15.8	43.1	29.8	20.8	18.1	23.9	--	--	15.1
Fe (ppm)	--	--	--	--	--	8.0	--	--	--	--	--	--	--	--
Solids Analysis (mmoles/gm)														
Ca	--	--	--	--	--	--	8.78	8.08	7.77	7.37	7.35	--	7.02	--
Mg	--	--	--	--	--	--	0.125	0.114	0.093	0.093	0.092	--	0.085	--
Na	--	--	--	--	--	--	0.277	0.350	0.303	0.721	0.522	--	0.954	--
Fe	--	--	--	--	--	--	--	--	--	--	--	--	--	--
SO_3	--	--	--	--	--	--	1.86	4.04	5.13	5.46	6.01	--	6.53	--
SO_4	--	--	--	--	--	--	0.31	0.52	0.52	0.68	0.68	--	0.79	--
CO_3	--	--	--	--	--	--	--	--	--	--	--	--	--	--

---

[a] Not available

[b] Flow on level control

[c] Flow on gravity control

Reproduced from best available copy.

TABLE D-2

STEAM COMPOSITION - MARCH 26, 1981

	Gas Streams					Feed Streams			
Stream Number	2	--	3	5	4	8	16	40	15
Location	Inlet Flue Gas	Scrubber Exit	Absorber Exit	Reheater Air	Stack Gas	Limestone Feed	Soda Ash Feed	Wash Water	Make-up Water
Temperature (°F)	283	132	117	--	110	--	--	--	--
Flow Rate (dscfm or gpm* or lbs/min**)	34,450	--	--	6,500	40,950	12.0**	7.0**	14.0*	0.0*
SO_2 (ppm)	2,050	--	95	--	80	--	--	--	--
O_2 (vol.% dry)	NA[a]	--	--	21.0	NA	--	--	--	--
Pressure Drop ("WG)	-0.4	6.7	1.9	--	--	--	--	--	--
Composition									
Percent Suspended Solids	--	--	--	--	--	99.94	99.90	--	--
Solids Analysis (mmoles/gm)									
Ca	--	--	--	--	--	9.62	--	--	--
Mg	--	--	--	--	--	0.163	--	--	--
Na	--	--	--	--	--	--	18.82	--	--
Fe	--	--	--	--	--	0.004	--	--	--
CO_3	--	--	--	--	--	9.78	9.41	--	--

	Process Liquor and Slurry Streams													
Stream Number	6	23	22	24	26	35	--	--	--	--	9	10	13	11
Location	Absorber Feed	Tray Feed	Absorber Recycle	Absorber Bleed	Scrubber Recycle	Scrubber Bleed	T-1 Bleed	T-2 Bleed	T-3 Bleed	T-4 Bleed	T-5 Bleed	Thickener Underflow	Filter Cake	Thickener Overflow
Temperature (°F)	NA[a]	NA[a]	118	118	120	120	120	120	119.5	119	119	--	--	NA[a]
Flow Rate (gpm)	140	140	2,560	LC[b]	1,160	128	GF[c]	GF[c]	GF[c]	GF[c]	LC[b]	--	--	LC[b]
Composition														
Percent Suspended Solids	--	--	--	--	--	--	1.23	1.13	1.24	1.26	1.51	17.9	33.9	0.150
Liquid Analysis (M)														
pH	6.26	--	--	6.16	--	5.92	5.94	5.97	6.00	6.01	6.07	--	--	6.07
HSO_3^-	0.565	--	--	0.638	--	0.950	0.838	0.788	0.790	0.780	0.750	--	--	0.675
$SO_3^=$	0.460	--	--	0.444	--	0.320	0.368	0.406	0.409	0.418	0.424	--	--	0.382
$SO_4^=$	0.99	--	--	1.023	--	1.10	1.08	1.070	1.080	1.080	1.070	--	--	0.980
HCO_3^-	0.013	--	--	0.012	--	0.004	0.008	0.016	0.012	0.012	0.012	--	--	0.009
H_2SO_3	0.025	--	--	0.025	--	0.008	0.017	0.033	0.024	0.024	0.024	--	--	0.017
Cl^-	--	--	--	--	--	0.087	--	--	--	--	--	--	--	0.0796
Mg^{++} (ppm)	673.0	--	--	708.0	--	780.0	778.0	785.0	768.0	828.0	800.0	--	--	658.0
Ca^{++} (ppm)	15.1	--	--	17.1	--	15.7	28.1	28.6	31.6	30.4	26.3	--	--	15.7
Fe (ppm)	6.0	--	--	6.0	--	8.0	6.0	6.0	6.0	6.0	5.0	--	--	5.0
Solids Analysis (mmoles/gm)														
Ca	--	--	--	--	--	--	8.21	7.77	7.71	7.45	6.89	--	6.66	--
Mg	--	--	--	--	--	--	0.077	0.067	0.071	0.062	0.062	--	0.046	--
Na	--	--	--	--	--	--	0.22	0.27	0.41	0.33	1.13	--	1.43	--
Fe	--	--	--	--	--	--	--	--	--	--	--	--	--	--
SO_3	--	--	--	--	--	--	3.78	4.68	5.19	5.89	6.00	--	6.77	--
SO_4	--	--	--	--	--	--	0.78	0.75	0.83	0.79	0.98	--	0.99	--
CO_3	--	--	--	--	--	--	4.41	3.06	2.27	1.30	0.79	--	0.06	--

---

[a] Not available

[b] Flow on level control

[c] Flow on gravity control

TABLE D-3

STREAM COMPOSITION - MARCH 27, 1981

	Gas Streams					Feed Streams			
Stream Number	2	--	3	5	4	8	16	40	15
Location	Inlet Flue Gas	Scrubber Exit	Absorber Exit	Reheater Air	Stack Gas	Limestone Feed	Soda Ash Feed	Wash Water	Make-up H_2O
Temperature °F	273	129	111	--	108	--	--	--	--
Flow Rate (dscfm or gpm* or lbs/min**)	48,830	--	--	6,640	55,470	26.0**	10.0**	15.0*	0.0*
SO_2 (ppm dry)	2,575	--	113.6	--	100.0	--	--	--	--
O_2 (vol.% dry)	6.7	--	--	21.0	8.4	--	--	--	--
Pressure Drop ("WG)	-0.4	6.6	1.9	--	--	--	--	--	--
Composition									
Percent Suspended Solids	--	--	--	--	--	99.98	99.90	--	--
Solids Analysis (mmoles/gm)									
Ca	--	--	--	--	--	9.66	--	--	--
Mg	--	--	--	--	--	0.137	--	--	--
Na	--	--	--	--	--	--	18.82	--	--
Fe	--	--	--	--	--	0.008	--	--	--
CO_3	--	--	--	--	--	9.80	9.41	--	--

	Process Liquor and Slurry Streams													
Stream Number	6	23	22	24	26	35	--	--	--	--	9	10	13	11
Location	Absorber Feed	Tray Feed	Absorber Recycle	Absorber Bleed	Scrubber Recycle	Scrubber Bleed	T-1 Bleed	T-2 Bleed	T-3 Bleed	T-4 Bleed	T-5 Bleed	Thickener Underflow	Filter Cake	Thickener Overflow
Temperature °F	91	NA[a]	114	114	118	118	118.0	118.0	117.5	117.5	117.0	--	--	91
Flow Rate (gpm)	150	150	2,521	LC[b]	1,127	140	GF[c]	GF[c]	GF[c]	GF[c]	LC[b]	--	--	LC[b]
Composition														
Percent Suspended Solids	--	--	--	--	--	--	3.08	3.14	3.40	3.62	3.50	12.50	36.94	1.91
Liquid Analysis (M)														
pH	6.27	--	--	6.07	--	5.63	5.94	6.08	6.12	6.16	6.18	--	--	6.12
HSO_3^-	0.493	--	--	0.613	--	1.03	0.759	0.655	0.647	0.633	0.607	--	--	0.603
$SO_3^=$	0.557	--	--	0.427	--	0.343	0.400	0.425	0.440	0.497	0.513	--	--	0.467
$SO_4^=$	0.94	--	--	0.95	--	1.02	1.00	1.00	1.03	1.05	1.07	--	--	0.93
HCO_3^-	0.013	--	--	0.013	--	0.03	0.03	0.007	0.01	0.013	0.017	--	--	0.013
H_2CO_3	0.026	--	--	0.026	--	0.05	0.06	0.013	0.02	0.027	0.033	--	--	0.026
Cl^-	--	--	--	--	--	0.080	--	--	--	--	--	--	--	0.081
Mg^{++} (ppm)	738.0	--	--	698.0	--	730.0	730.0	748.0	763.0	800.0	760.0	--	--	760.0
Ca^{++} (ppm)	18.6	--	--	14.5	--	21.3	41.8	47.5	32.9	41.9	40.0	--	--	16.5
Fe (ppm)	5.0	--	--	3.0	--	2.0	2.0	1.0	1.0	1.0	1.0	--	--	5.0
Solids Analysis (mmoles/gm)														
Ca	--	--	--	--	--	--	7.93	7.56	7.49	7.45	7.43	--	5.75	--
Mg	--	--	--	--	--	--	0.072	0.056	0.061	0.055	0.056	--	0.292	--
Na	--	--	--	--	--	--	0.22	0.35	0.33	0.59	0.28	--	3.49	--
SO_3	--	--	--	--	--	--	4.75	5.87	6.09	6.37	6.47	--	6.04	--
SO_4	--	--	--	--	--	--	0.45	0.95	0.73	0.79	0.76	--	1.22	--
CO_3	--	--	--	--	--	--	3.08	1.49	1.13	0.68	0.65	--	0.04	--

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[a] Not available

[b] Flow on level control

[c] Flow on gravity control

Fit Page Numbering

APPENDIX E

EQUIPMENT PROBLEMS

10 Lines
to End of
Typing Area

EQUIPMENT AND MAINTENANCE PROBLEMS

Section	Problem	Frequency*	Cause	Action/Solution
General	Piping leaks	H	Corrosion of mild steel piping, erosion of rubber lining and consequent corrosion of mild steel piping and flange leaks	Piping repaired or replaced with appropriate material
	Pluggage of pipes with foreign material	H	Inadequate cleanout of all process piping and equipment prior to startup and accidental dropping of foreign matter into process equipment	Foreign matter cleaned out
	Pluggage of pipes	H	Crystallization of solids and salts, lack of heat tracing of all lines and equipment	Cleaned out and heat traced whenever possible
	SO_2 meters not operable	-	Not adequately recommissioned prior to startup	Recommissioning was completed during operation
	pH meters not reliable	H	Erosion and corrosion of the sensor, electrical problems and inadequate recalibration	Operated plant with portable pH meter as control
	Control room instruments operate and indicate poorly	H	Insufficient recommissioning of the instrumentation and lack of available spare parts	No action
	Control panel instrumentation overheating	H	Air conditioning in control room was inadequate to cool the equipment or failed	Added a second air conditioner in the control room
	Instrument air compressor/air supply failure	L	Open piping	Piping was secured
	Purge water failures	H	Incorrectly adjusted purge water valves	Instrumentation of cleaned and purge water adjusted
	Purge water failures	H	Inadequate supply pressure due to large demand by filter sprays	Filter water sprays were conve to river wate
	Total electrical failure	L	Problems were external to plant	No action
Water Supply	River water totalizer pluggage	H	Totalizer filled with debris from the river	Cleaned out the totalizer
	Well water line leak	H	---	Repaired
	Well water pump failures	H	Overheated motor	Replaced motor

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*O = Once
L = Low Frequency
H = High Frequency

EQUIPMENT AND MAINTENANCE PROBLEMS

Section	Problem	Frequency*	Cause	Action/Solution
SO_2 Injection	Inoperable at time of plant startup	-	Delivered inoperable with inadequate documentation for on-site repairs	Requested documentation from leasor
	Low nitrogen pressure on the SO_2 tank	O	Leak in the nitrogen piping	Repaired
	Difficulty in maintaining constant SO_2 injection rate	H	Poor instrumentation set up for the vaporizer	Adjusted time constants and proportionality constants. Manual operation
	Leak in SO_2 injection pipe	O	---	Repaired
	SO_2 analyzer failure	L	Air conditioning in the analyzer shack failed	Repaired
F.D. Fan	Rust carryover into the scrubber	H	Natural buildup of rust in the steel ductwork	Flushed solution from the scrubber to the pond
	Excessive vibration	L	Impeller out of balance	Balanced impeller. Repaired bearings and motor coupling
Scrubber	Gas leak at inlet expansion joint	O	---	Repaired
	Plumb bob would not move	O	Air motor does not operate	Repaired
	Scrubber pump failures	H	Rubber lining failures, motor failures, leaking packing, electrical switch gear failures, flange leaks	Repaired
	Scrubber piping failure	O	Rubber lining failed at an elbow in the recirculation piping	Replaced with stainless steel piping
	Scrubber pump valve failures	H	Inappropriate valves for the service	No action
	Scrubber lining failures	L	---	Repaired
Scrubber Instrumentation	Automatic level control failure	H	Setting of time constants and proportionality constants, valve failures, level transmitter failures	Level control was accomplished manually
	Leaking through control valves	H	Inappropriate valves for the service	No action in most cases

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EQUIPMENT AND MAINTENANCE PROBLEMS

Section	Problem	Frequency*	Cause	Action/Solution
Absorber	Automatic level control failures	H	Same as above	Controlled manually
Reactors	Failure of the bottom of Reactor 1	O	Erosion due to buildup of small "pebbles"	Cleaned out and repaired the bottom of the tank
	Failure of agitator of Reactor 4	O	Tripped out on overload	Adjusted automatic trip mechanism
	Pluggage in Reactor 5	O	---	Cleaned out
	Overflow pipe pluggage	L	Crystallization of solids	Overflow pipes were cleaned or replaced
Reactor Instrumentation	Automatic level control of Reactor 5 failure	L	Failure of the control valve	Repaired
	pH meters	H	Not functional	No action
Thickener	Rake failure	O	Lift mechanism failure caused overtravel which broke the rake gearbox housing	Repaired
	Rake jammed	O	Stopping the rake to allow maximum settling	Restarted using air/liquor sparger to free the rake
	Thickener underflow pump seal failures	H	Loss of pressure on the seal water line when the filter was in operation	Piped the filter wash water to the river water system
	Underflow piping pluggage	H	Crystallization of the solids due to cold weather, settling	Cleaned out the piping and heat traced as much as possible
	Underflow pumps	H	Inadequately sized to maintain flow to filter	Operated both pumps in parallel or diluted the underflow
	Thickener bottom failure	O	Inadequate check-out during recommissioning	Thickener drained and repaired. Grouting of bottom
	Uneven flow over weir	H	Flow between weir and its support rather than over the weir	Caulked the weir
Filter	Poor control of vacuum and blowback	H	Incorrect adjustment of the automatic valve	Adjusted and refaced the automatic valve when necessary
	Hole in the filter cloth	H	---	Changed the cloth
	Crack in filter housing	L	---	Repaired

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EQUIPMENT AND MAINTENANCE PROBLEMS

Section	Problem	Frequency*	Cause	Action/Solution
Filter Cont.	Crack in internal piping	L	---	Repaired
	Ineffective scraper	H	Misaligned and inappropriate design	Adjusted and eventually removed the scraper
	Filtrate receiver pump failure	O	Hole in pump housing	Repaired
Filter Instrumentation	Pressure Gauge Failure	H	---	Repaired
Reheater	Piping leaks	H	---	Repaired
	Pilot light failure	O	Inadequately sized pressure reducing valve	Replaced
	Flame failure	O	Plugged oil nozzle	Cleaned
	Flame failure	O	Poor alignment of the flame sensor	Realigned
	Inadequate reheat	H	Undersized dilution air fan	Discontinued reheater operation
Pipe to Pond	Pluggage of pipe	H	Not flushing the line after each use	Cleaned out
	Leaks	H	---	Repaired
Limestone Feeder	Readout failure	H	Integrated circuit board failure	Replaced
	Limestone chute jamming	H	Moisture above Reactor 1	Cleaned out
Soda Ash System	Soda ash feeder failure	H	Circuit board failure	Replaced
	Soda ash jam in the chute	H	Moisture above the soda ash tank	Cleaned the chute
	Soda ash tank agitator failure	O	Coupling failed	Repaired
	Piping leaks	H	Inappropriate specification of materials for the pipe from the thickener hold tank to the soda ash tank	Replaced with appropriate material
	Soda ash feeder conveyor failure	H	Buildup of soda ash around the conveyor	Cleaned out
	Soda ash feeder readout in the control room failure	H	Circuit board failure	Repaired

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